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FUEL GAS ENVIRONMENTAL IMPACT: PHASE REPORT

Fred L. Robson, et al

United Technologies Research Center

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FUEL GAS
ENVIRONMENTAL IMPACT:
PHASE REPORT

by

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16. ABSTRACT The report gives results of an evaluation of the technical and economic feasibility of: (1) Lurgi-type fixed-bed gasifiers and BCR-type entrained-flow gasifiers in combination with low- and high-temperature fuel gas cleanup systems; (2) advanced technology combined-cycle power systems; and (3) integrated gasification systems, cleanup processes, and power systems. Processes and systems considered were those using technology both currently available for power station configurations which the contractor judged could appear in commercial applications in the 1975-78 time frame (first generation systems) and potentially applicable in the 1980-decade time period (second generation systems). The results indicate that high-temperature cleanup systems have the potential of improving the efficiency and reducing the capital costs of integrated gasification systems.

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ABSTRACT

A program was carried out to evaluate the technical and economic feasibility of: (1) Lurgi-type fixed-bed gasifiers and BCR-type entrained-flow gasifiers in combination with low- and high-temperature fuel gas cleanup systems, (2) advanced technology combined-cycle power systems, (3) integrated gasification systems, cleanup processes and power systems. Processes and systems considered were those using technology currently available for power station configurations which the Contractor judged could appear in commercial applications in the 1975-1978 time frame (first-generation systems) and those using technology potentially applicable in the 1980-decade time period (second-generation systems). The objective of this analytical study of fuel gas environmental impact is the definition of combinations of: (1) fossil fuel gasification systems, (2) low- and high-temperature fuel gas cleanup processes, and (3) advanced-cycle power systems for central power stations that appear to result in the lowest practicable emissions of air, water, and solid pollutants consistent with the environmental constraints, while producing low-cost electrical power.

The method of analysis is based upon the systems approach in which the technical and economic characteristics of the overall integrated gasification, cleanup and power system are evaluated as a whole. A Contractor-owned digital computer program was utilized to define the performance of the system from coal in to kilowatts out. The modular approach to analysis by this unique analytical tool permits wide flexibility in fuel process configurations and power cycle arrangement. However, lack of substantial data on gasifier operation limited the approach to design point calculations.

The analyses indicate that high-temperature cleanup systems have the potential of improving the efficiency and reducing the capital costs of integrated gasification systems. However, unacceptable emission levels for NO_x could result with some gasifier types. The use of commercially available low-temperature cleanup systems results in lower efficiencies and higher costs, but would still allow generation of electrical power at costs competitive with conventional steam stations with stack gas cleanup while having sulfur emissions which are below regulations for conventional stations.

This report was submitted in partial fulfillment of Project Nos. 21 ADD 27 and 104, Contract No. 68-02-1099, by the Research Center of the United Technologies Corporation, under the sponsorship of the Environmental Protection Agency. This report covers work performed between July 1, 1973 and November 1, 1974.

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LIST OF CONVERSION FACTORS

Btu x 0.252 = Kcal

ft x 0.3048 = m

in. x 25.4 = mm

F subtract 32 x 0.555 = C

lb x 0.453 = Kg

lb x 0.453 = Kg

scf (@ 60 F & 30 in. Hg) x 0.0284 = m^3 (@ 15.5 C & 762 mm Hg)

Btu/scf x 8.88 = Kcal/m^3 (@ 15.5 C & 762 mm Hg)

lb/10^6 Btu x 1.798 = Kg/10^6 Kcal

ton x 1.104 = metric ton

CONCLUSIONS

1. First-generation coal gasifiers and commercially available low-temperature cleanup processes used in conjunction with combined-cycle power systems have the potential of producing electrical power at costs competitive with conventional coal-fired steam plants with stack gas cleanup. These integrated gasification/power systems would also have lower emissions of SO_2, NO_x, and particulates, reduced thermal pollution, and would provide recovered sulfur in the elemental form.

2. High-temperature cleanup processes used in conjunction with combined-cycle power systems have the potential of providing significant advances in efficiency and lower capital costs which could result in the production of electrical power at costs 20% or more lower than conventional coal-fired steam plants with stack gas cleanup.

3. Emissions of NO_2 from power systems using high-temperature cleanup processes could be unacceptable. These higher NO_x emissions result from: (a) higher combustion temperatures due to the higher temperature fuel; and (b) fuel-bound nitrogen compounds formed in some types of gasifiers which would pass directly through the high-temperature cleanup process. It may be possible to mitigate the NO_x formation due to temperature effects by careful combustor design or modifications. At this time, there appears no acceptable method of removing fuel-bound nitrogen at high temperature.

4. Gasifiers operating at temperatures beyond those considered in this study (1800 F) would tend to break down fuel-bound nitrogen to N_2 and H_2 and thus could reduce the fuel-bound nitrogen carryover and subsequently the NO_x emissions from the power system.

5. Overall system performance improvements resulting from advancing gas turbine inlet temperatures 400 F (second-generation power systems) are comparable to those improvements associated with the change from low-temperature to high-temperature cleanup processes.

6. Steam power systems using gasification/cleanup processes appear competitive with conventional coal-fired steam systems with stack gas cleanup only when second-generation gasifiers and high-temperature cleanup processes are used in conjunction with a let-down turbine to recover compressor power. These systems would be subject to NO_x emission problems similar to those of the combined-cycle system.

7. Particulate carryover from the gasifier may pass through the high-temperature cleanup process with little or no reduction. Thus, both the environment and the turbine may suffer. However, some high-temperature particulate removal

devices, such as the Aerodyne-type high energy cyclone, appear to be nearer commercialization than do the cleanup processes.

8. The high-temperature iron oxide process, which is nearer commercialization than other high-temperature processes, requires a larger amount of clean fuel to convert the SO_2 to elemental sulfur. Other forms of ultimate sulfur recovery would not be as energy intensive.

RECOMMENDATIONS

The overall attractiveness of the integrated gasification/cleanup/combined-cycle systems even when using essentially present-day technologies lead to the recommendation that this system should be demonstrated in a large enough scale (~ 100 Mw or larger) to obtain experience on systems which could be commercialized at an early date. Prior to, or parallel with, this expensive effort (about $300 million for two programs) several other less ambitious but essential programs are recommended:

1. The potential benefits of the high-temperature cleanup system are great enough to warrant extensive development effort to bring these systems to commercialization at an early date. Programs to verify physical configuration, materials of construction, and sulfur capture ratios at pressures and temperatures typical of combined-cycle operations (e.g. 300-400 psi, 1500-1700 F) should be carried out in a fuel gas stream, preferably from a gasifier or from a synthesized gas source.

2. Methods of removing nitrogen compounds, mainly ammonia, from high-temperature fuel gas streams should be investigated. In addition, studies should be initiated on the mechanism of ammonia formation during gasification with the objective of determining operating conditions consistent with minimum ammonia production. Alternatively, gasifiers operating at temperatures at which ammonia breaks down to nitrogen and hydrogen should be considered.

3. In the off gas from a gasifier, the actual carryover of materials potentially harmful to the gas turbine; e.g., particulates, alkiline metals, vanadium, etc. should be measured. Only when these data are available can performance and cost estimates be made with relative confidence.

4. Programs leading to the commercialization of high-temperature particulate removal devices should be continued or initiated at the easrliest possible time. Removal efficiencies of 99% down to 2μ are estimated to be required for adequate turbine life and to meet environmental constraints.

5. The present study was constrained to specific gasifier operating points due to a lack of adequate data. The development of a simple model of the gasifier would allow preliminary estimates to be made of variations in overall system performance due to changes in gasifier operation.

6. The potential emission of thermal NO_x, (NO_x resulting from the combustion process, not from fuel nitrogen) from high-temperature, high-pressure ratio gas turbines should receive further consideration. The preliminary model of NO_x formation presented in this study should be further developed, if possible, to take into consideration more extensive current and future data on the combustion of low-Btu fuel in gas turbine combustors.

INTRODUCTION

There are many possible coal gasification/advanced-cycle power system combinations that may be integrated into future power generating installations. Before committing large resources to any specific combination, an assessment must be made of the technical and economic characteristics and the environmental intrusion from the coal conversion process, the fuel gas cleanup system, and the power cycle. Towards this objective an analytical program was undertaken by UTRC and the Foster Wheeler Energy Corporation to define combinations of coal gasification systems, low- and high-temperature fuel gas cleanup processes, and advanced power cycles that appear to result in the lowest practicable emissions of air, water and solid pollutants while producing low-cost electric power from central power stations.

When initiated, this study was a part of the EPA Control Systems Laboratory plan to evaluate methods of reducing or eliminating emissions from coal-fired central generating stations. Additional impetus was given to the EPA program by the 1973 October War and the resulting embargo on mid-east petroleum imports. Subsequent examinations of this nation's energy supplies and identification of the goals of Project Independence have spotlighted coal as the "fuel of the future." Thus, the prior work done by the EPA and its predecessor, the National Air Pollution Control Administration (NAPCA) on pre- and post-combustion cleanup of coal has attained even greater importance.

The energy crunch with its accompanying conservation measures has reduced somewhat the projected growth of the electric utility industry. Nonetheless, even conservative estimates(1) of the need for electrical energy in 1980 indicate that about 625 GW will be required, of which 425 GW will be fossil fueled. By 1995, it is estimated that 1350 GW will be installed, 610 GW of which would be fossil fueled.

With all the projected additional installed capacity of fossil-fueled power systems, the already difficult problem of utility-caused sulfur dioxide (SO_2), particulate, and nitrogen oxides (NO_x) pollution could become intolerable. There are three alternative solutions open for utility consideration: (1) use of naturally occurring clean fuels, (2) treatment of the powerplant exhausts to remove pollutants, and (3) treatment of dirty fuels before combustion to remove harmful pollutants. Because of the very high costs and demand for the naturally occurring clean fuels by other segments of the economy having little or no alternative, widespread use of such fuels will not be attractive to a major portion of the utility industry. The second alternative, stack gas cleanup, shows promise as a method of meeting the near-term (1975) EPA standards for SO_2 and these systems, some with the help of special particulate removal devices, will meet the standards for particulate emission levels for 1975. Currently, however, no commercial-scale stack gas cleanup system has been demonstrated that also removes significant amounts of NO_x, nor have removal efficiencies much beyond those required for 1975 standards been demonstrated conclusively. The third alternative, fuel pretreatment, can in several of its forms such as hydrodesulfurization of residual oils or solution of coal, meet near-term goals for sulfur and particulates. However, only gasification followed by some type of sulfur cleanup process has the promise of attaining sulfur levels significantly below those required by 1975 standards.

In order to offset some of the inefficiencies in converting coal to a clean fuel, an advanced-cycle power system of high efficiency could be utilized. The use of gasified coal for fueling advanced-cycle power systems is not a new idea. As far back as the mid-1960's Steinkohlen-Elektrizität AG (STEAG) and Lurgi Gesellschaft für Wärme und Chemotechnik mbH proposed(2) a combined-cycle plant using Lurgi gasifiers to fuel a pressurized boiler which would exhaust into a gas turbine. The original concept did not involve a fuel gas desulfurization step although one has subsequently been added. Widespread attention to the concept of the integrated gasifier/cleanup/power system was brought to focus by the publication(3) in December 1970 of the United Aircraft Research Laboratories/Burns and Roe, Inc./FMC report "Technological and Economic Feasibility of Advanced Power Cycles and Methods of Producing Non-Polluting Fuels for Utility Power Stations." The major conclusion of this report was that advanced technology available in the aircraft gas turbine industry could be applied to advanced-cycle COmbined Gas And Steam (COGAS) power systems which would allow the use of costlier gasified and desulfurized fuel in power stations which would generate essentially pollution-free power at costs competitive with or less than conventional steam stations with no pollution control.

This study was based upon NAPCA-prescribed power systems of 1000-Mw nominal size and technology judged best available, i.e., technology, which could appear if given immediate and necessary research and development funding. In the years following the publication of this report, it is apparent that the gasified coal-fired COGAS system, while recognized as the most attractive future method of generating clean power, has not received funding in the amounts and within the time required to match the study projections. Additionally, subsequent effort(4) at UTRC has indicated that performance benefits might be obtained through the use of high-temperature fuel gas cleanup systems, but the use of such systems could result in increased NO_x production over systems using low-temperature fuel gas cleanup processes. The technology level used in these later studies has been that judged to be attainable based upon currently available funding and also upon marketing considerations; i.e., projected acceptance of COGAS systems and its effect on the development of large, advanced-technology, industrial turbines.

In order to put into perspective the technical and economic characteristics of these advanced power systems based upon the current assessment of technological development and also to identify the potential environmental intrusion of the systems, a study was conducted to evaluate combinations of gasification, cleanup processes, and power cycle configurations capable of producing low-cost, pollution free electric power from coal. The results of that study are presented in this report. Section 1 contains the definition of the operating characteristics of a fixed-bed gasifier (U.S. Bureau of Mines) and a two-stage entrained-flow gasifier (Bituminous Coal Research) and a review of low- and high-temperature fuel gas cleanup processes. In Section 2, the operating characteristics of advanced power systems that could be used with the gasification and cleanup systems are identified. Preliminary integration efforts leading to the selection of systems for detailed evaluation are also given in Section 2. The results of the foregoing analysis were used to select applicable cleanup processes for use in integrated gasification/cleanup/power systems. Low- and high-temperature cleanup processes for each gasifier type were selected for detailed technical and economic evaluation. Detailed schematics and flow sheets were prepared. This work is presented in Section 3. In Section 4, the performance and cost of the integrated gasification cleanup and power systems is presented. As a basis for comparison, performance and cost of a direct coal fired steam system is also given. Capital costs of the selected systems are tabulated and resultant power costs compared. Lastly, environmental goals were selected and comparison of the emissions from the various systems with these goals were carried out. In addition, the status of the technology required to meet the various environmental goals was identified and recommendations made for further effort in these areas. This work is reported in Section 5.

The report contained herein covers the first three phases of a four phase study. The results presented are thus preliminary in nature, subject to reevaluation during the fourth phase. Many of the areas of analysis identified in the recommendations will be carried out during the fourth phase.

SECTION I

GASIFIER AND CLEANUP SYSTEM PROCESS REVIEW

This section presents a survey of low- and high-temperature fuel gas cleanup processes which could provide essentially sulfur-free fuel to advanced-cycle electric power generating stations. These cleanup systems would be used in conjunction with two generations* of coal gasifiers: the first-generation gasifier would have off gases containing condensible tars as typified by the fixed-bed type (e.g., Lurgi or McDowell-Wellman); the second-generation gasifier would be of the entrained flow type having off gases without condensibles (e.g., BCR-two-stage).

COAL GASIFICATION SYSTEMS

Production of low-calorific fuel gas from coal is achieved by gasification with air and steam at elevated temperatures. The overall gasification process is endothermic primarily due to the steam-carbon reaction which requires about 5000 Btu/lb of carbon:

$$C + H_2O \rightarrow CO + H_2 \qquad (1)$$

This heat requirement is usually satisfied by burning a portion of the coal feed with air.

Gasification systems are commonly classified into three categories according to the characteristics of the gasifier coal bed; i.e., fixed, fluidized, or entrained. Generally speaking, fixed-bed systems operate with countercurrent flow of coal and gas at temperatures below the ash fusion point.

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*First-generation systems are those using current technology which the Contractor judges could appear in the 1975-1978 time frame. Second-generation systems are those using technology judged by the Contractor to be potentially applicable for commercial configurations in the 1980-decade.

Consequently, these systems are characterized by relatively low gasification rates and substantial tar formation. High-temperature gasification under ash slagging conditions affords higher gasification rates together with reduced reactor volume and essentially a tar-free producer gas. Co-current entrained flow gasifiers typically operate under the latter conditions. However, at the higher temperatures, the fraction of the coal heating value represented by sensible heat of the product gas is substantially greater than for fixed-bed gasifiers. To achieve maximum thermal efficiency, this sensible heat must be used effectively. It is this problem of effective utilization that forms a major portion of this study.

For use in advanced-cycle power systems the producer gas is required at elevated pressure. Since the volume of the producer gas is roughly twice that of the air required for gasification, it is normally advantageous to employ pressure gasification rather than product gas compression to obtain a given delivery pressure. Moreover, the specific gasification rate, gas produced per unit of reactor volume, is greater at higher pressures.

First-generation technology for pressure gasification of coal is limited to the fixed bed-type gasifier. Second-generation technology will probably evolve from current developmental efforts on entrained-flow and fluid-bed gasifiers. For the purpose of this study, the two-stage entrained flow-type gasifier was selected as representing advanced coal gasification technology. Typical operating characteristics of fixed-bed and entrained-flow gasifiers are given in Table 1.

The two prominent differences in the performance of these gasifiers are directly related to their operating temperature:

1. The fixed-bed system operates at 900 F to 1100 F and has relatively low carbon conversion and a large amount of by-product tar is formed.

2. The entrained-flow gasifier operates at 1800 F and above and has a ratio of sensible heat of product gas to coal heating value twice as great as the fixed-bed system. No condensibles are formed.

The presence of tar in the off gas of the fixed-bed gasifier complicates downstream heat recovery and subsequent gas purification. For low-temperature purification of the raw gas, direct quenching is the preferred method, thereby avoiding heat exchanger fouling, but the level for sensible heat recovery is degraded. The recovered tar product could be recycled to the gasifier to increase the overall carbon gasification. If a high-temperature purification

Table 1

COMPARISON OF GASIFICATION SYSTEMS

Gasifier type:	Fixed-Bed	Two-Stage Entrained-Flow
Temperature, F	1000	1800
Pressure, psia	500	500
Coal	Western Kentucky	Illinois No. 6
Air/Coal, lb/lb	2.69	3.42
Steam/Coal, lb/lb	0.349	0.567
% Carbon Gasified*	84	99
Producer Gas		
SCF/lb Coal	58.8	74.6
Vol. %		
N_2	47.61	47.70
CO	20.55	16.74
CO_2	5.88	8.84
H_2	13.83	11.98
CH_4	2.76	3.14
H_2S	0.60	0.46
COS	0.10	0.10
NH_3	0.25	0.38
H_2O	8.42	10.66
	100.00	100.00
Tar, lb/lb coal	0.11	0.0
HHV, Btu/scf (tar free)	139	125
Sensible Heat/Coal HV, %	12.2	24.6

*No tar recycle

system is integrated with the fixed-bed gasifier such that no tar is allowed to condense, then the tar represents a component in the delivered producer gas. This may present a problem from a pollution standpoint in that the tarry material typically contains 1.5-3.0 percent sulfur and about 1.0 percent nitrogen which, during combustion, will convert to SO_2 and NO_x.

The large sensible heat content of the gas from an entrained-flow gasifier makes it imperative, from an efficiency standpoint, to use heat recovery techniques in conjunction with low-temperature purification. Fortunately, the absence of condensible tars facilitates indirect heat exchange and affords the possibility for regenerative heating of the fuel gas sent to the power production system. Because of the high effluent temperature, the entrained flow gasifier is particularly amenable for integration with high temperature purification systems operating in the range of 1400-1700 F.

Gasifier Operating Data

Descriptions and operating data on the fixed-bed (Lurgi) and entrained-flow (BCR) gasifiers were to be Government-furnished information. For the analytical program carried out during this study, the following gasifier data were required: coal, steam, and air feed rates and conditions, operating temperatures and pressures, fuel gas output temperature, pressure, and composition, size limitations on the gasifiers, capital and operating costs, potential emissions of pollutants such as ash, coal, tar derivatives, or thermal discharge to the environment. Fuel gas constituents such as sulfur compounds, alkaline metals, chlorine, and particulate matter which could be harmful to the environment or to the operation of downstream system components were to be quantitatively identified.

Unfortunately, the Government-furnished information was not available in a timely fashion. The study program was therefore based on gasifier operating data collected by Foster Wheeler Corporation and UTRC from other sources. The information so obtained, although the best available, was single point data which necessitated extrapolation to conditions applicable to the subject study. Furthermore, there was a definite lack of information on the gasification system costs as well as on the pollutants discharged from the gasifier.

Bureau of Mines Stirred-Bed Gasifier - The Bureau of Mines (BOM) stirred-bed coal gasifier was selected as the gasification system for the study of first-generation powerplants. This selection was made on the basis of the availability of operating data and the similarity in operational mode to the Lurgi system which is the only present-day high-pressure gasifier in commercial operation.

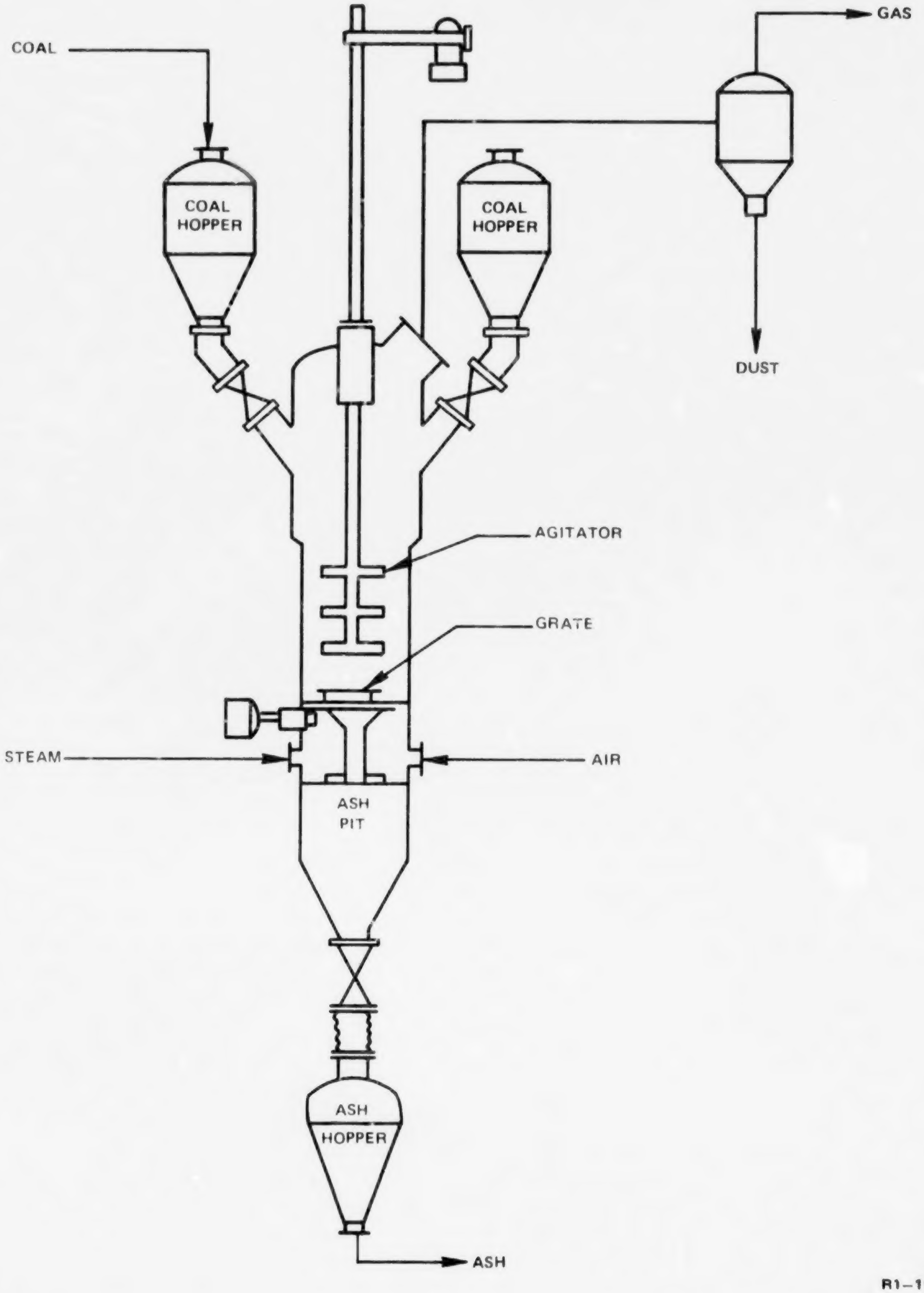
A schematic diagram of the BOM gasifier is shown in Figure 1. Although this system is not commercially available, a one ton/hour pilot unit is in operation in the Morgantown Energy Research Center and a commercial-sized unit is in the design stage. Because of the rabble arm agitator, the gasifier is capable of processing both caking and noncaking coals.

In operation(5), coal screened to 1 x 2 inches is fed intermittently to the top of the bed through lock hoppers pressurized with inert gas. Air and steam are admitted below the rotating grate and the coal bed moves downward countercurrent to the ascending stream of hot gases produced from partial combustion of coal. The steam rate is adjusted to control the bed temperature below the ash fusion point to avoid clinker formation. Ash is discharged by eccentric rotation of the grate and is removed through a pressurized lock hopper. Producer gas exiting from the top of the converter is partially cleaned of entrained dust in a cyclone separator.

The Bureau of Mines supplied a summary of raw data for 35 runs made during the 1973 operation of the pilot gasifier. These runs used four Eastern caking coals as feedstock: Illinois No. 6 (1.3 percent sulfur), Loveridge (2.9 percent sulfur), Upper Freeport (3.8 percent sulfur), and Western Kentucky No. 9 (3.9 percent sulfur). These data were evaluated with the object of providing a basis for the study of the first-generation gasifier. Eight runs showed an acceptable material balance (±5 percent) and, arbitrarily, the highest pressure acceptable run made on the high-sulfur Western Kentucky coal (Run 55-3) was selected. The data from this run was adjusted to close on elemental balances, particularly sulfur, which showed only 65 percent recovery. In addition, the heat balance was adjusted to reduce the large heat rejection to jacket cooling water which represented 16 percent of the feed coal heating value. Considering that a commercial-scale gasifier would have a much lower surface to volume ratio, a 3 percent heat rejection to jacket cooling was taken as a reasonable value. The adjusted heat and material balance is given in Table 2 for a Western Kentucky coal. The cold gas efficiency of this process, defined as the higher heating value of the tar-free producer gas divided by the higher heating value of the coal feed, is 72.5 percent. In a commercial operation where the tar components are condensed out, it would be advantageous from an overall fuel utilization viewpoint to recycle the tar to the gasifier. The heat and material balances for the BOM gasifier were therefore revised to take into consideration recycle of by-product tar. Since the pilot gasifier has not been operating in this fashion, the resulting gasifier yields shown in Table 3 represent a judicious extrapolation to commercial scale

FIG. 1

BUREAU OF MINES STIRRED – BED GASIFIER


GAS
COAL
COAL HOPPER
COAL HOPPER
DUST
AGITATOR
GRATE
STEAM
AIR
ASH PIT
ASH HOPPER
ASH

R1–19–1

Table 2

BUREAU OF MINES GASIFIER

GASIFICATION OF WESTERN KENTUCKY COAL

Feed Coal Analysis:

	C	H	S	O	N	ASH	H_2O
Wt. %	64.47	5.48	3.90	4.18	1.58	15.44	4.95

H.V. = 11,450 Btu/lb

Gasifier Operation:

Temperature, F	1000
Pressure, psia	135
Air (150 F, 185 psia), lb/lb Coal	2.733
Steam (1165 F, 145 psia), lb/lb Coal	0.349

Gasifier Products:

*Gas, scf/lb Coal	59.21
Tar, lb/lb Coal	0.108
Dust, lb/lb Coal	0.016
Ash, lb/lb Coal	0.166

Gas Analysis:

	N_2	CO	CO_2	H_2	CH_4	H_2S	H_2O
Vol. %	48.26	20.16	6.06	13.49	2.74	0.71	8.58

M.W. = 25.049 HHV = 140.3 Btu/scf

Tar Analysis:

	C	H	S	N+O
Wt. %	82.1	7.6	1.5	8.8

H.V. = 16,000 Btu/lb

*Dust and Tar Free

Table 3

BUREAU OF MINES GASIFIER

GASIFICATION OF WESTERN KENTUCKY COAL WITH TAR RECYCLE

Feed Coal Analysis:

	C	H	S	O	N	ASH	H_2O
Wt. %	64.47	5.48	3.90	4.18	1.58	15.44	4.95

H.V. = 11,450 Btu/lb

Gasifier Operation:

Temperature, F	1000
Pressure, psia	135
Air (150 F, 185 psia) lb/lb Coal	3.187
Steam (1165 F, 145 psia) lb/lb Coal	0.405

Gasifier Products:

*Gas, scf/lb Coal	68.92
Dust, lb/lb Coal	0.016
Ash, lb/lb Coal	0.166

Gas Analysis:

	N_2	CO	CO_2	H_2	CH_4	H_2S	H_2O
Vol. %	48.24	20.15	6.06	13.52	2.74	0.64	8.65

M.W. = 24.926 HHV = 139.9 Btu/scf

*Dust and Tar Free

operations.* This extrapolation was made by assuming the same distribution of gasified carbon as the base case data, using the same air to steam ratio, and maintaining the same approach to the water gas shift equilibrium. The cold gas efficiency of the gasification process with tar recycle is thereby increased to 84.2 percent.

The BOM data did not indicate the presence of any ammonia or carbonyl sulfide in the raw product gas, although these compounds would be expected to be formed at the gasification conditions. In the subsequent use of these data, the material balances were adjusted to reflect the presence of NH_3 and COS by assuming that all of the nitrogen in the coal feed was converted to NH_3 (conservative) and that H_2S and COS were present in the off-gas in a volumetric ratio of 10 to 1.0.

BCR Entrained-Flow Gasifier - The two-stage entrained flow gasifier developed by Bituminous Coal Research, Inc. (BCR) was selected as the second-generation gasification system for this study. An oxygen-blown, 120 ton/day pilot plant to produce 2.5 MMSCFD of SNG is under construction at Homer City, Pennsylvania. Design of an air-blown version is currently underway at the Foster Wheeler Corporation.

A schematic diagram of the BCR gasifier with its auxiliaries is shown in Fig. 2.(6) Run-of-the-mine coal is crushed, dried, and pulverized until 70 percent passes through a 200 mesh screen. The pulverized coal is metered from the feed hopper into hot transport gas recycled from the gas purification section, and then fed into the upper stage of the gasifier. In this stage, the coal reacts with synthesis gas from the lower stage together with steam producing methane, carbon monoxide, hydrogen, and unreacted char. The gases leave the upper stage at around 1800 F.

Entrained residual char is removed from the gas by cyclone separators and recycled via superheated steam to the lower stage of the gasifier. The char then reacts with steam and air at about 2800 F to form synthesis gas and molten slag. The hot synthesis gas flows to the upper stage for reaction with coal as described above. Molten slag collects and drains from the bottom of the lower stage into the slag pot where it is water quenched.

Overall, the gasifier reactions are endothermic and the process heat requirement is supplied by combustion of char with air. The air rate is regulated

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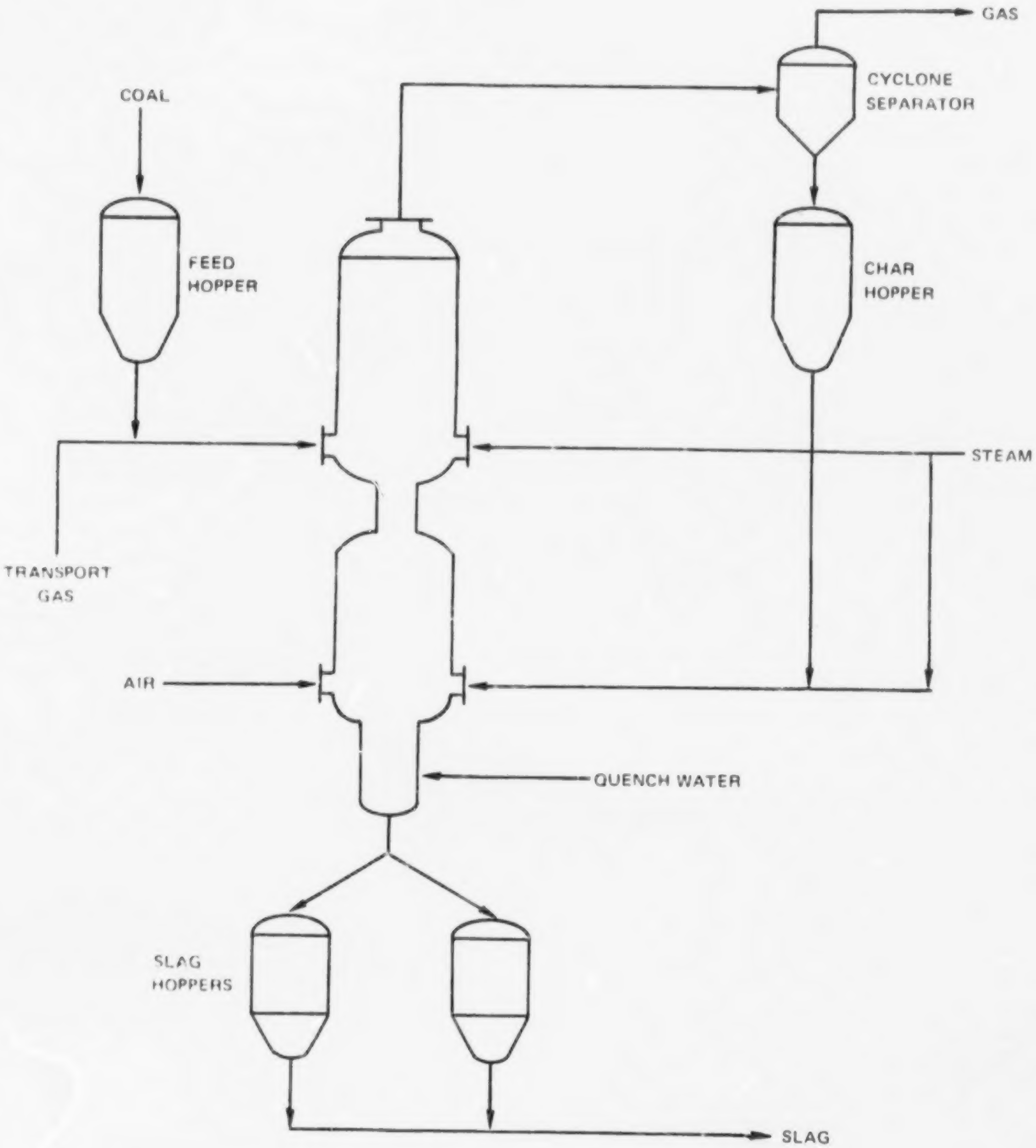
*Commercial systems using Lurgi gasifiers typically recycle the tar or utilize the tars as a binder for briquetting coal fines into sizes required for use in the gasifier.

FIG. 2

BCR ENTRAINED – FLOW GASIFIER


GAS
CYCLONE SEPARATOR
COAL
FEED HOPPER
CHAR HOPPER
STEAM
TRANSPORT GAS
AIR
QUENCH WATER
SLAG HOPPERS
SLAG

R1–19–3

to maintain the operating temperature in the upper stage while the lower stage temperature is controlled by steam addition. Temperature in the lower stage is fairly critical since too high a temperature will damage the refractory and too low a temperature will cause the ash slag to freeze and accumulate.

Operating data for the two-stage BCR gasifier was supplied by Gilbert Associates, Inc.(7) who were acting as project monitor on an OCR contract for the design and construction of a 1200 ton/day developmental gasifier system. A summary of the data made available on the production of a low-Btu fuel gas from Illinois No. 6 coal is given in Table 4. This information is judged to represent the expected performance of the BCR gasifier although it remains to be confirmed via actual operating experience.

The calculated cold gas efficiency for this gasification system is 83.3 percent. Content of particulates, chlorine, alkaline metal vapor, and trace components in the product gas was not available. It is expected that the high gasifier operating temperature with sufficient residence time would preclude the formation of coal tars, naphthas, and phenols resulting in a relatively clean product gas compared to the fixed-bed gasifier. Coal nitrogen was assumed to form ammonia.

FUEL GAS CLEANUP PROCESSES

Fuel gas cleanup systems have been divided into two categories; those requiring substantial cooling of the dirty fuel gas to 250 F or below (low-temperature systems) and those requiring little or no cooling of the dirty gas prior to cleaning (high-temperature systems).

Low-Temperature Cleanup Systems

Low-temperature processes for desulfurizing raw producer gas are commercially available and have been widely used for natural gas sweetening and treating synthesis gas in the chemical process industry; e.g., the manufacture of ammonia, methanol, oxo alcohols. These systems normally operate below 250 F and are commonly classified into the following categories:

Chemical Solvent Processes
Physical Solvent Processes
Direct Conversion Processes
Dry Bed Processes

Table 4

BCR ENTRAINED-FLOW GASIFIER

GASIFICATION OF ILLINOIS NO. 6 COAL

Feed Coal Analysis:

	C	H	S	O	N	ASH	H_2O
Wt. %	67.4	5.1	3.8	9.6	1.2	8.7	4.2

HHV = 12,200 Btu/lb

Gasifier Operation:

Temperature, F	1800
Pressure, psia	455
Transport Gas (300 F, 555 psia), scf/lb Coal	6.560
Air (800 F, 515 psia), lb/lb Coal	3.088
Steam (950 F, 915 psia), lb/lb Coal	0.567

Gasifier Products:

Gas, scf/lb Coal	77.27
Slag, lb/lb Coal	0.087

Product Gas Analysis:

	N_2	CO	CO_2	H_2	CH_4	H_2S	COS	NH_3	H_2O
Vol. %	45.76	18.13	8.31	13.07	3.60	0.49	0.10	0.39	10.15

M.W. = 24.519 HHV, Btu/scf = 141.9

Transport Gas Analysis:

	N_2	CO	CO_2	H_2	CH_4	NH_3	H_2O
Vol. %	49.15	19.58	6.49	14.10	3.91	0.08	6.69

M.W. = 24.240 HHV, Btu/scf = 147.6

In Table 5 are listed the gas treating processes available for removing sulfur compounds, particularly hydrogen sulfide, from raw producer gas. This list was compiled from a literature survey(8-11) and, while it is believed to represent the major technology in this area, it is not necessarily all inclusive. Figure 3 illustrates the process scheme for a typical absorption-stripping process for low-temperature acid gas removal.

Chemical Solvent Processes - Chemical solvent processes employ aqueous solutions of organic and/or inorganic agents which are capable of forming "complexes" with the acid gas components, notably H_2S and CO_2, present in the raw gas stream. The absorption solution is regenerated by decomposing the "complex" at elevated temperature thereby releasing the acid gases for subsequent recovery and recycling the solution for further absorption. These processes are essentially insensitive to the partial pressure of acid gases in the feed and generally exhibit little or no selective absorption of H_2S over carbon dioxide. The chemical processes may be sub-divided into those using amine scrubbing solutions and those based on alkali scrubbing solutions.

The principal reactions involved in gas sweetening with amine solutions (10-30 percent weight) may be represented as:

$$RNH_2 + H_2S \rightarrow RNH_3\ HS \qquad (2)$$

$$RNH_2 + CO_2 + H_2O \rightarrow RNH_3\ HCO_3 \qquad (3)$$

Monoethanolamine (MEA) will easily reduce the H_2S content below 4 ppm; however, it is not considered selective, even though the rate of CO_2 absorption is less than for H_2S. The principal disadvantage of MEA is that it will react with COS and CS_2 forming nonregenerable compounds. Diethanolamine (DEA) will not react with these contaminants and is favored for service where COS and CS_2 are likely to be present. Like MEA, DEA solutions are not selective for H_2S, and, will seldom reduce the H_2S content below 100 ppm. Tertiary amines, such as triethanolamine and methyl-diethanolamine, while not as reactive as the other amines, have the advantage of being selective towards H_2S removal. The tertiary amines are two to four times more costly and find little application in industrial gas sweetening.

The alkali scrubbing system may be represented by the following chemical reactions:

$$M_2\ CO_3 + H_2S \rightarrow MHS + MH\ CO_3 \qquad (4)$$

$$M_2\ CO_3 + CO_2 + H_2O \rightarrow 2\ MH\ CO_3 \qquad (5)$$

Table 5

LOW-TEMPERATURE DESULFURIZATION PROCESSES

PROCESS	SORBENT	TEMPERATURE F	SELECTIVE H_2S REMOVAL	STATUS
Chemical Solvent Processes				
MEA	Monoethanolamine	80-120	No	Commercial
SNPA:DEA	Diethanolamine	100-130	No	Commercial
TEA	Triethanolamine	100-150	Yes	Not Commercially Important
MDEA	Methyl-diethanolamine	100-150	Yes	Not Commercially Important
Econamine	Diglycolamine	80-100	No	Commercial
ADIP	Di-Isopropanolamine	80-100	No	Commercial
Alkazid	Potassium Dimethyl Amino Acetate	70-120	Yes	Commercial
Hot Potash	Potassium Carbonate Solution	200-250	Partial	Obsolete
Catacarb	Activated Potassium Carbonate Solution	150-250	Partial	Commercial

Table 5 - Continued

PROCESS	SORBENT	TEMPERATURE F	SELECTIVE H_2S REMOVAL	STATUS
Benfield	Activated Potassium Carbonate Solution	150-250	Partial	Commercial
Seaboard	Sodium Carbonate Solution	80-100	Partial	Obsolete
Vacuum Carbonate	Sodium Carbonate Solution	80-100	Partial	Obsolete
TPP	Tri-Potassium Phosphate Solution	80-120	Partial	Commercial
Sodium Phenolate	Sodium Hydroxide + Phenol Solution	80-100	Partial	Obsolete
Aqua Ammonia	Aqueous Ammonia Solution	60-120	Partial	Commercial

Table 5 - Continued

PROCESS	SORBENT	TEMPERATURE F	SELECTIVE H_2S REMOVAL	STATUS
Physical Solvent Processes				
Sulfinol	Sulfolane + Di-Isopropanolamino	80-120	Possibly	Commercial
Selexol	Polyethylene Glycol Ether	20-80	Yes	Commercial
Fluor Solvent	Propylene Carbonate	40-80	Yes	Commercial
Purison	N-Methyl Pyrrolidone	70-100	Yes	Commercial
Rectisol	Methanol	< Zero	Yes	Commercial
Estasolvan	Tri-N-Butyl Phosphate	80-120	Yes	Not Commercially Important
Water Wash	Water	70-100	Yes	Not Commercially Important

Table 5 - Continued

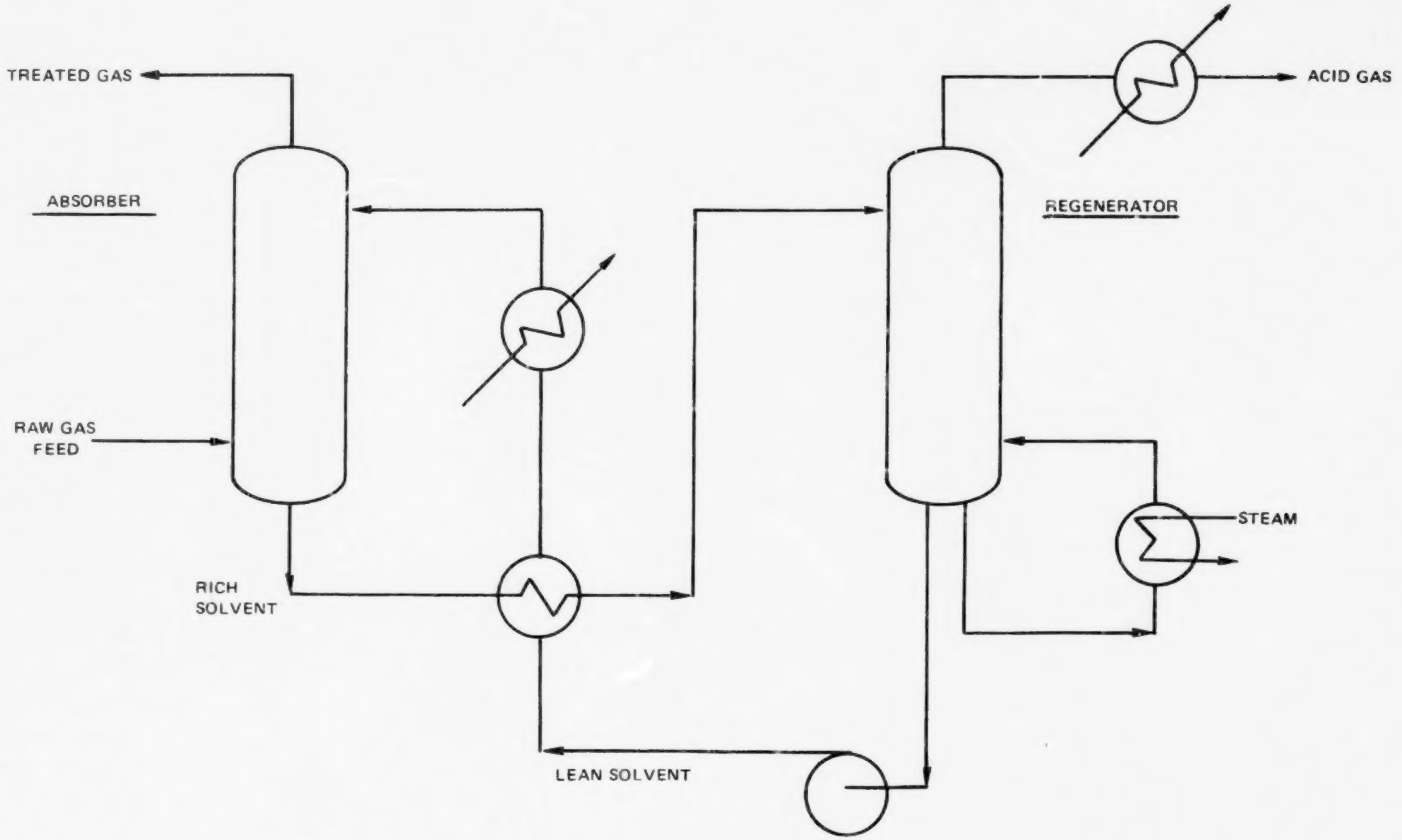
PROCESS	SORBENT	TEMPERATURE F	SELECTIVE H_2S REMOVAL	STATUS
Direct Conversion Processes				
Stretford	Sodium Carbonate + Anthraquinone Sulfonic Acid	--	Yes	Commercial
Takahax	Sodium Carbonate + Nathoquinone Sulfonic Acid	--	Yes	Commercial (outside U.S.)
Giammarco-Vetrocoke	Arsenic Activated Potassium Carbonate	100-300	Yes	Commercial (outside U.S.)
Thylox	Arsenic Trioxide + Sodium Carbonate	--	Yes	Commercial (outside U.S.)
Ferrox	Iron Oxide + Sodium Carbonate	--	Yes	Commercial (coutside U.S.)
Manchester	Iron Oxide + Sodium Carbonate	--	Yes	Commercial (outside U.S.)
Perox	Aqueous Ammonia + Hydroquinone	--	Yes	Commercial (outside U.S.)

Table 5 - Continued

PROCESS	SORBENT	TEMPERATURE F	SELECTIVE H_2S REMOVAL	STATUS
Lacy-Keller	Proprietary	--	Yes	Not Commercially Important
Townsend	Triethylene Glycol	100-250	Yes	Not Commercial
IFP	Proprietary	200-300	Yes	Not Commercial
Deal Process	Aqueous Sulfolane	--	Yes	Not Commercial
Nalco Process	Proprietary	--	Yes	Not Commercial
Dry Bed Processes				
Haines	Zeolites (+SO_2)	70-100	Yes	Not Commercially Important
Molecular Sieve	Molecular Sieve (adsorption)	70-100	Yes	Commercial
Iron Sponge	Hydrated Ferric Oxide	70-100	Yes	Commercial
Activated Carbon	Carbon	70-100	Yes	Commercial

FIG. 3

TYPICAL LOW – TEMPERATURE ACID GAS REMOVAL UNIT


TREATED GAS
ABSORBER
RAW GAS FEED
RICH SOLVENT
LEAN SOLVENT
REGENERATOR
ACID GAS
STEAM

R1–19–2

The earlier processes, such as Seaboard and Vacuum Carbonate, were based on dilute solutions of sodium carbonate (3-4 percent weight) and were capable of removing 80-90 percent of the H_2S. Regeneration in the Seaboard process was by air resulting in a dilute acid gas stream while the Vacuum Carbonate system used vacuum regeneration with steam. These processes were superseded by the hot potassium carbonate system. In the "hot pot" processes, an aqueous solution of 25-35 weight percent K_2CO_3 is used to absorb acid gases at temperatures in the range of 200-250 F. With low H_2S/CO_2 ratios, the process is capable of sweetening the gas to 5 ppm. A degree of selective H_2S absorption over CO_2 can be achieved by taking advantage of the relatively slow rate for CO_2 absorption. In addition to removing H_2S and CO_2, the process can remove COS and CS_2 by hydrolysis of these components to CO_2 and H_2S. The Catacarb and Benfield processes are improved versions of the Bureau of Mines "hot pot" systems insofar as they employ activators to increase the rate of absorption thereby decreasing the required solution circulation rate. Disadvantages of the hot potassium carbonate systems are the relatively high steam consumption for regeneration, required operating pressure above 300 psi, and the inability to remove mercaptans.

The tripotassium phosphate process was developed by the Shell Oil Company specifically for H_2S removal via the reaction:

$$K_3PO_4 + H_2S \rightarrow K_2HPO_4 + KHS \qquad (6)$$

The nonvolatility of the agent, its nonreactivity with COS and CS_2, and partial selectivity toward H_2S in the presence of CO_2 gives it certain advantages over the amine systems. However, when operated for high H_2S selectivity, the process only gives about 90 percent removal efficiencies, and conversely, with high H_2S removal the steam consumption becomes excessive due to CO_2 absorption.

Physical Solvent Processes - Physical solvent processes all use organic solvents to remove acid gases by physical absorption, rather than chemical reaction, which is directly proportional to the partial pressure of the acid gas components. These processes are most applicable to high-pressure gas treating where appreciable quantities of sour gases are present. After absorption, the "loaded" solvent is regenerated by heat and/or pressure reduction giving a concentrated stream of H_2S plus CO_2 and a recyclable lean solvent. Due to the higher solubility to H_2S in these organic solvents, selective absorption of H_2S over CO_2 can be achieved. In general, these processes have two major disadvantages; the solvents have a great affinity for absorbing heavy hydrocarbons (C_5+) which contaminate the gas stream fed to sulfur recovery units, and the solvents are quite expensive so that large solvent losses cannot be tolerated.

As a group, these processes were developed for bulk removal of acid gases but, for low H_2S concentration, they are capable of giving a sweetened gas having less than 5 ppm H_2S. In order to maximimize the solubility of acid gases and minimize solvent loss through vaporization, the processes are generally operated at or below ambient temperature. In addition to removing H_2S and CO_2, the solvent processes are all capable of removing COS, CS_2 and mercaptans without solvent degradation as well as dehydrating the gas to a low dew point. The low heats of solution for acid gases result in appreciably lower steam requirements for solvent regeneration compared to the chemical solvents.

The Sulfinol process is unique in that it combines the characteristics of a solvent process and an amine process. The physical absorbent, Sulfolane, gives high acid gas loadings at high acid gas partial pressures, giving it bulk removal capacity and the chemical absorbent, DIPA* reduces residual acid gases to very low values. However, the presence of the chemical solvent reduces the H_2S selective absorption for this system compared to the straight solvent processes.

Direct Conversion Processes - Two types of processes fall into this category:

a. those based on oxidation reduction reactions, and

b. those based on the stoichiometric reaction of H_2S with SO_2 in the presence of a solvent.

The first type involves the absorption of H_2S in alkaline solutions containing oxygen carriers. The H_2S is subsequently oxidized to elemental sulfur by air fed to the regenerator where the sulfur product is flotated and collected at the regenerated solution interface as a froth. Processes of this type are in common use in Europe for removal of H_2S and sulfur recovery from manufactured gases and coke-oven gas. The Ferrox and Manchester processes employ a suspension of iron oxide in an aqueous solution of sodium carbonate to absorb H_2S. With multistage absorption, essentially complete removal of H_2S is obtainable; however, the product is of low quality due to salt contamination and chemical replacement costs are high. Both the Thylox and Giammarco Vetrocoke(12-13) processes use alkaline solutions of arsenates and are capable of reducing the H_2S to less than 1 ppm. Partial removal of COS, CS_2, and mercaptans is also possible. Again the sulfur product is contaminated and the use of arsenates makes these processes potentially hazardous. The Stretford and Takahax processes are similar in that alkaline solutions of quinone sulfonic acids are employed. The addition of vanadium salts increase the rate of oxidation of hydrosulfide to sulfur resulting in higher solution loadings. Close to 100% H_2S removal is claimed for these processes along with high purity (99%) sulfur product. However, substantial amounts of thiosulfates are formed resulting in sludge deposition and corresponding chemical makeup.

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*di-isopropanolamine

Generally speaking, the low solution loadings exhibited by this group of processes make them uneconomical for treating large very sour gas streams. They are best suited for sour gases containing less than 1.0% H_2S with sulfur production under 20 tons/day. These processes, as well as those in the following group are almost totally selective for H_2S removal.

In the second group of direct conversion processes are those in which H_2S is absorbed in a solvent and converted to elemental sulfur by the Claus type reaction with SO_2.

$$2\ H_2S + SO_2 \rightarrow 3\ S + H_2O \qquad (7)$$

The Townsend process uses an aqueous solution of an organic solvent, such as triethylene glycol, to sweeten the gas, dehydrate the gas, and convert H_2S to elemental sulfur. A portion of the product sulfur is burned to SO_2 which is absorbed by fresh solvent and the SO_2 rich solvent is used to contact the sour gas. The IFP, Nalco, and Deal processes operate in a similar manner employing other solvents. While a high purity (99.7%) sulfur product is claimed, none of these processes have been commercialized.

Dry Bed Processes - These sweetening processes are based on absorption of acid gases by a fixed bed of solid absorbent. Due to their low absorbent loading, they are applicable to gases containing low concentration of H_2S and mercaptans, perhaps less than 500 ppm. These processes can be subdivided into the iron oxide processes and the various molecular sieve processes.

The iron oxide or dry box process is one of the oldest processes known for removing sulfur compounds from industrial gases. In the iron sponge system, wood shavings impregnated with hydrated ferric oxide are used to absorb H_2S:

$$2\ Fe_2O_3 + 6\ H_2S \rightarrow 2\ Fe_2S_3 + 6\ H_2O \qquad (8)$$

Regeneration of the absorbent is carried out with air:

$$2\ Fe_2S_3 + 3O_2 \rightarrow 2\ Fe_2O_3 + 6S \qquad (9)$$

This process is best suited for small to medium gas volumes with low sulfur contents, otherwise the sponge bed life would be too short to be economical. The process is selective towards H_2S and mercaptans and will partially remove COS and CS_2. Sweetened gas of less than 5 ppm H_2S is easily obtained. However, sulfur recovery would not be economical when using the iron sponge system.

Molecular sieves can be tailor made to have pore sizes which will permit selective absorption of H_2S over CO_2. These processes are characterized by the various regeneration schemes employed; i.e., via hot combustion gases, hot SO_2 gas as in the Haines process, or hot air. In the latter two modes, elemental sulfur is produced via the oxidation of the absorbed H_2S. The sieve processes also appear to be economically attractive for small to medium gas volumes having low H_2S content. Additionally, for efficient H_2S removal, the raw sour gas should have a water content below 20 lb/MMSCF since water will also be absorbed by the molecular sieve structure.

Selection Considerations

Low-temperature desulfurization systems for application in low-Btu fuel gas plants will have to treat large volumes of sour gas, 500-1000 MMSCFD, having total sulfur content in the range of 0.4 to 1.0% (weight). In addition to H_2S, the raw gas will contain CO_2, COS, CS_2, probably mercaptans, cyanides and heavy hydrocarbons. Of the types of processes described above, it is evident that the liquid scrubbing processes, physical solvents and some chemical solvents, are the best suited. These processes are currently available and can easily reduce the sulfur content to 100 ppm, which, when combusted, would result in SO_2 emissions well within present EPA regulations for conventional steam stations. As such, these processes are capable of serving both first generation and second generation coal gasification plants.

High-Temperature Cleanup Systems

High-temperature systems for sulfur removal are not presently available in commercial scale although there are several in various stages of development. Most of the active work presently involves use of limestone and dolomites which have potential in the range of 1500-2000 F. Other systems receiving attention employ iron oxides, molten salts, and liquid metals. These systems operate by chemical reaction of the absorbent with sulfur compounds in the gas, forming the corresponding metal sulfides. The degree of desulfurization attainable depends on the chemical equilibria for the particular system at operating conditions. As with low-temperature processes, economics dictate that the sulfided absorbent be regenerated for reuse.

The only commercial experience with high-temperature desulfurization reported in the literature is that of the Frodingham Desulfurization Process.(14-15) This process employed fluidized beds of pulverized iron oxide operating at 650 F to 800 F. In the early 1960's, a commercial plant treating 32 MMCFD of crude coke oven gas containing 1.0% H_2S was operated at the Exeter Works of the South Western Gas Board. Essentially complete (99.9%) removal of H_2S was achieved with 90% removal of all organic sulfur compounds other than thiophene. The sulfided absorbent was regenerated with air at 1000-1100 F resulting in a SO_2 stream which was subsequently converted to sulfuric acid. Major difficulties were experienced in the solids handling system which produced fine oxide dust resulting in excessive losses of the absorbent. In addition, operation of the sulfuric acid plant was erratic due to low SO_2 concentrations in the regenerator off-gas.

Several processes currently under development which may prove commercially viable for use with second generation gasification systems are listed in Table 6.

Bureau of Mines Process(16-17). This process, under development at the Morgantown Energy Research Center, is based on a sintered absorbent consisting of a mixture of iron oxide (Fe_2O_3) and fly ash. This sorbent satisfied the primary requirements for high-temperature sulfur removal in that it is readily available and inexpensive, it has reasonable absorption capacity for sulfur, can be regenerated for repeated use, and is resistant to fusion and disintegration over the operating temperature range of 1000-1500 F. The absorbent is prepared by mixing iron oxide with "as received" fly ash to a total iron oxide content of about 35%. Iron oxide contents above 40% were unsatisfactory because the fusion temperature was lowered within the operating range. The mixture is extruded into 1/4" x 3/8" pellets and then sintered to develop the required hardness.

Absorption studies over the range of 1000-1500 F show sulfur capacities of 10 to 25% by weight, respectively, for dry simulated producer gas. The presence of water vapor reduces the capacity to 6-10% by weight but there is no evidence of loss of absorption effectiveness over 150 cycles of regenerations.

The reaction mechanism is chemisorption wherein hydrogen sulfide diffuses into the sorbent particle and reacts with iron oxide forming iron sulfide.

$$Fe_2O_3 + 3H_2S \rightarrow FeS + FeS_2 + 3H_2O \qquad (10)$$

$$FeO + H_2S \rightarrow FeS + H_2O \qquad (11)$$

Sulfided absorbent is regenerated with air at temperatures of 1000-1500 F producing an SO_2 containing off-gas and reusable Fe_2O_3. Since sulfur recovery in the elemental form is preferable from a pollution standpoint, the formation of SO_2 is a disadvantage to this process.

In accordance with the above absorption reactions, the residual sulfur content in the treated gas is governed by the chemical equilibrium relationships:

$$[H_2S] = [H_2O] / K_1^{1/3} \qquad (12)$$

$$[H_2S] = [H_2O] / K_2 \qquad (13)$$

Table 6

HIGH TEMPERATURE DESULFURIZATION PROCESSES

Process	Absorbent	Temperature, F	Status
Bureau of Mines	Sintered Iron Oxide + Fly Ash	1000-1500	Pilot
Babcock & Wilcox	Iron Oxide	700-1200	Experimental
Consolidation Coal (CONSOL)*	Half-Calcined Dolomite	1500-1800	Pilot
Air Products	Calcined Dolomite	1600-2000	Pilot
Battelle Northwest	Molten Carbonates	1100-1700	Pilot
IGT-Meissner	Proprietary	800	Unknown

*Now Conoco Coal Development Company

where [ ] = the gas phase mol fraction of the component and K_1, K_2 = equilibrium constants for reactions (10) and (11), respectively. K_1 and K_2 are given in Figs. 4 and 5. Removal of H_2S by this process is independent of operating pressure, and for a given gas composition, will decrease with increasing temperature. Iron oxide catalyzes the water gas shift reaction.

$$H_2O + CO \rightarrow H_2 + CO_2 \qquad (14)$$

so that the gas composition approaches equilibrium with respect to this reaction as it passes through the absorption bed. Equilibrium constants for the shift reaction are shown in Figure 6.

Data on the absorption of H_2S from actual producer gas(19) indicate that the equilibrium absorption closely corresponds to equation (13) above.

Although the absorption of other sulfur compounds on this sorbent have not been studied, thermodynamically, the removal of COS appears to be practical:

$$COS + FeO \rightarrow FeS + CO_2 \qquad (15)$$

where, $[COS] = [CO_2]/K_3$, K_3 = equilibrium constant.

Values for K_3 are given in Figure 7.

To date, this process has been operated on a pilot scale absorbing H_2S in the raw producer gas from a 1.0 ton/hour stirred-bed coal gasifier. The H_2S content was reduced from 0.6% to 150 ppm at 1100-1200 F and 120 psi. Tar present in the raw gas was not removed by the absorbent. Further work on the regeneration cycle is currently in progress. Commercial-scale plants would be based on the multiple fixed-bed principle alternating between absorption and regeneration cycles. Fluidized bed operation would not be considered due to potential attrition of the absorbent.

Consolidation Coal Process - This process(20) evolved as an adaption of the CO_2 Acceptor Process(21) for producing low-Btu fuel gas from coal and incorporates the use of a half-calcined dolomite acceptor for sulfur capture as studied by Squires and coworkers(22-25). Basically the process chemistry involves the following reaction:

$$[CaCO_3 \cdot MgO] + H_2S \rightarrow [CaS \cdot MgO] + H_2O + CO_2 \qquad (16)$$

for which the equilibrium H_2S concentration in the gas is given by

$$[H_2S] = [H_2O]\,[CO_2]\,P/K \qquad (17)$$

FIG. 4

EQUILIBRIUM CONSTANT FOR H_2S ABSORPTION BY IRON OXIDE

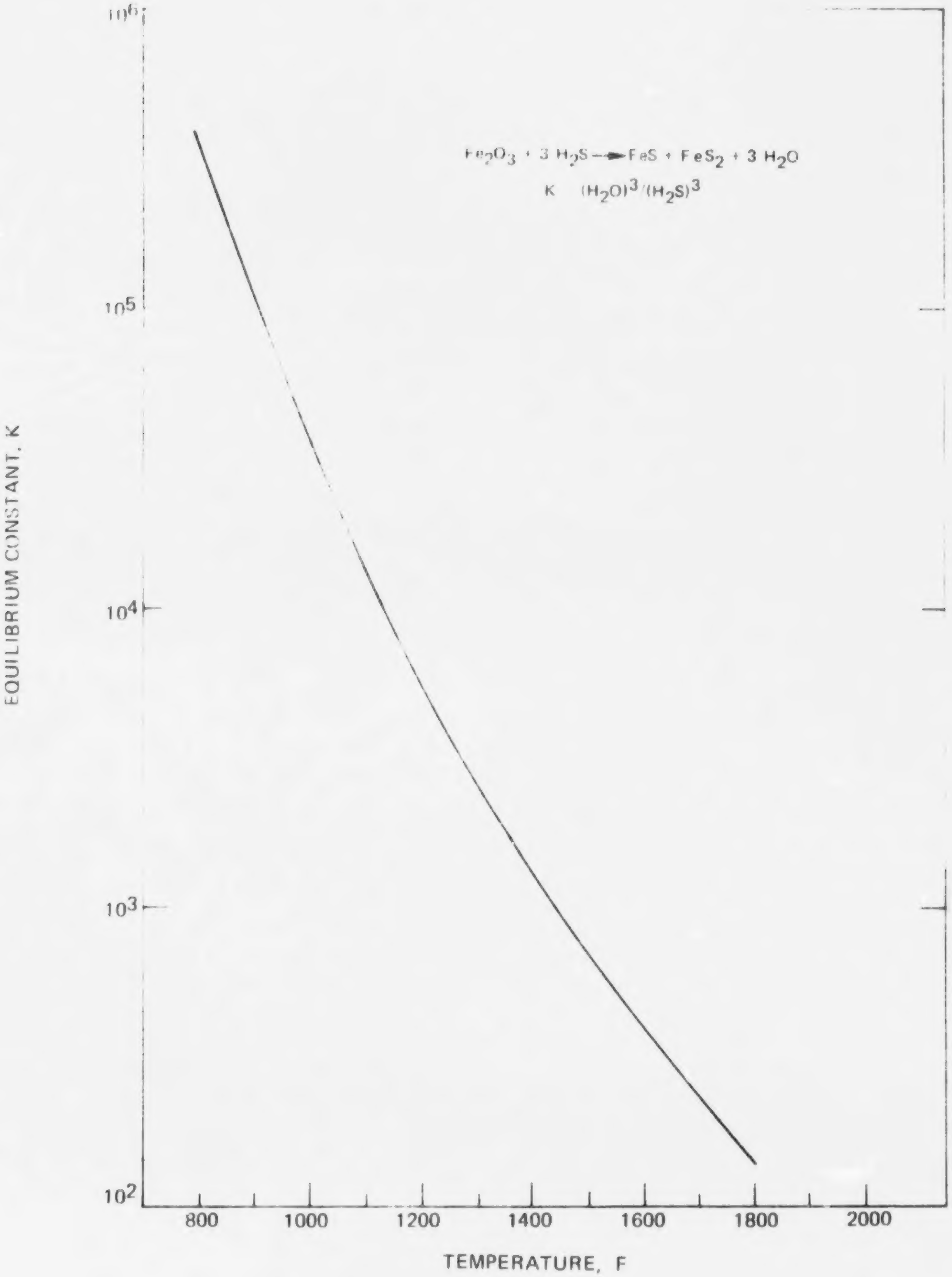

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FIG. 5

EQUILIBRIUM CONSTANT FOR H_2S ABSORPTION BY IRON OXIDE

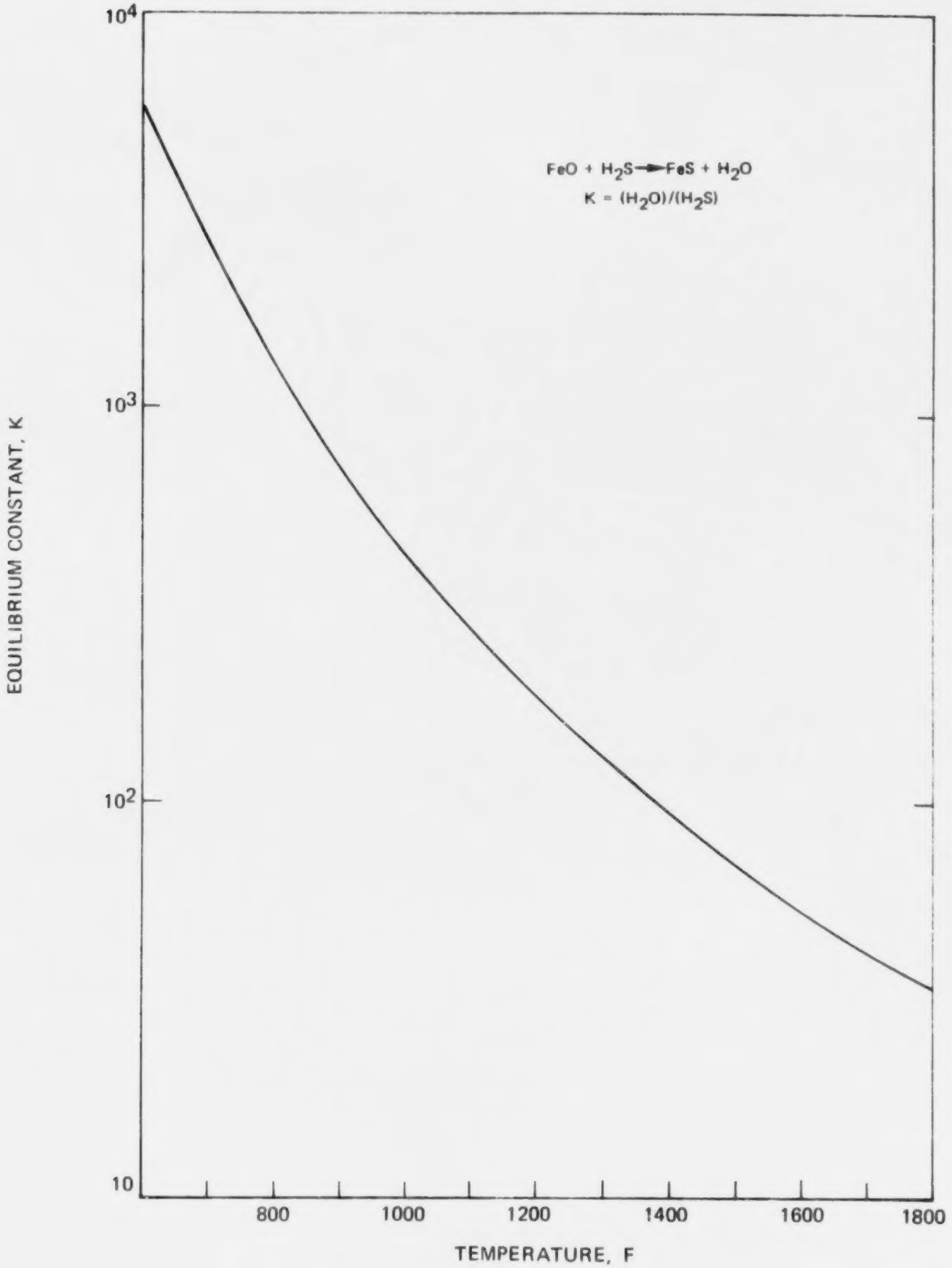

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FIG. 6

EQUILIBRIUM CONSTANT FOR WATER GAS SHIFT

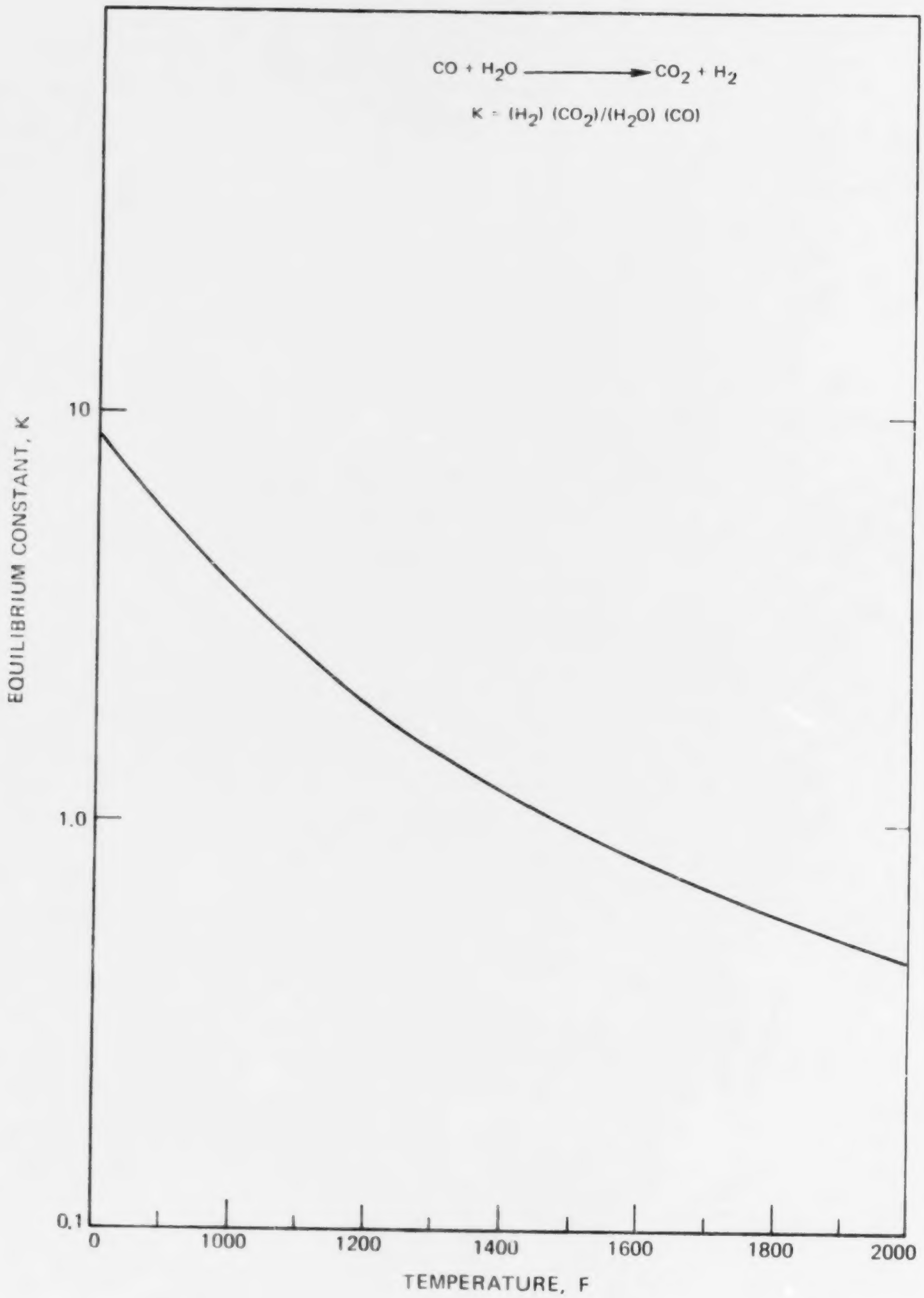
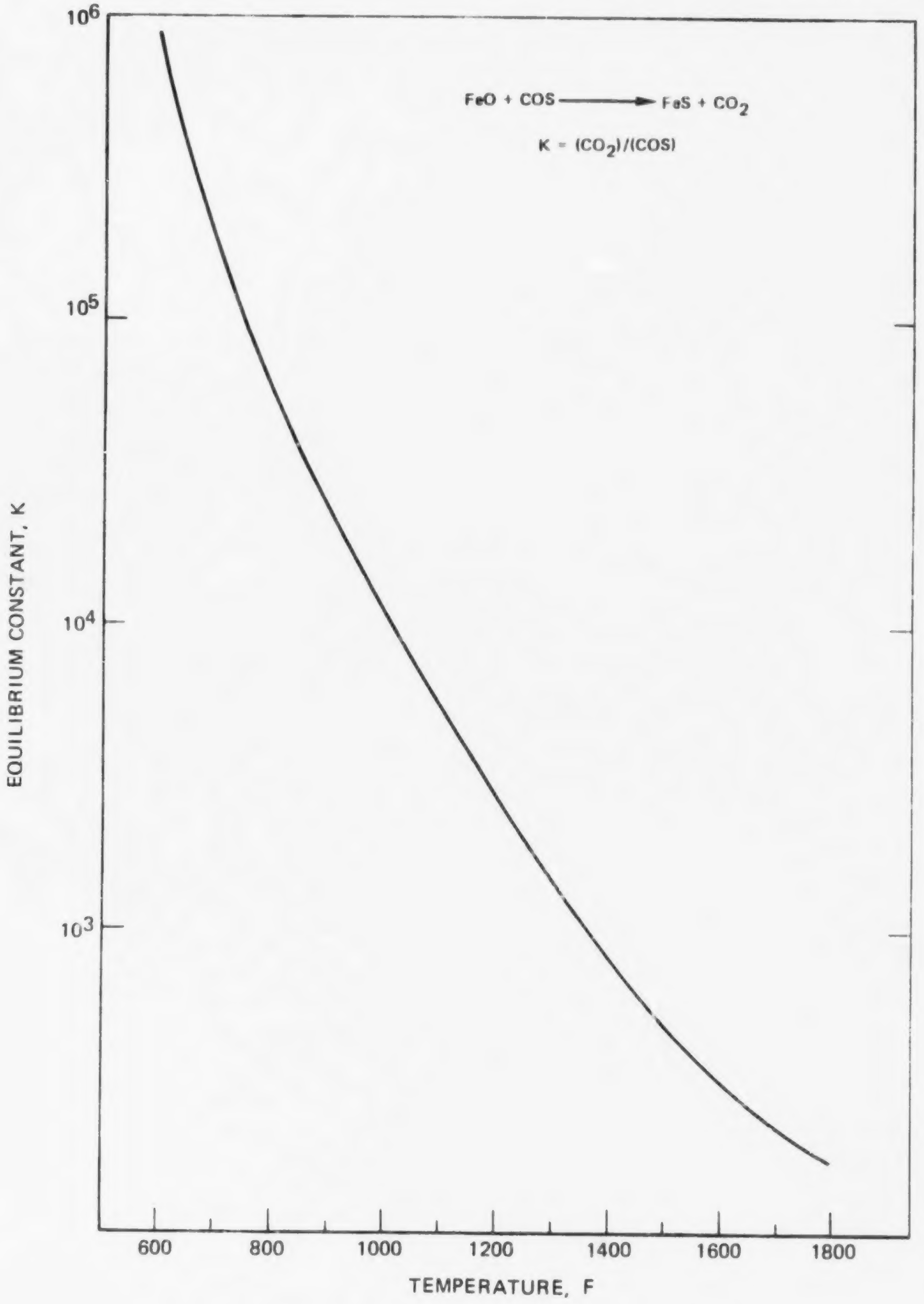


FIG. 7

EQUILIBRIUM CONSTANT FOR COS ABSORPTION BY IRON OXIDE



R1–19–23

where $[\ ]$ = mol fraction of component

P = total pressure, atm

K = equilibrium constant

From Fig. 8, showing the variation of K with temperature, it is obvious that the removal of H_2S increases with temperature for a given gas composition. Also, for a given gas composition and temperature, the removal efficiency is inversely proportional to the operating pressure.

A maximum operating temperature for this process is imposed by the partial pressure of carbon dioxide in the gas phase; i.e. the temperature cannot exceed that at which the CO_2 partial pressure is equal to the decomposition pressure for $CaCO_3$ via the following reaction:

$$Ca\,CO_3 \rightarrow CaO + CO_2 \tag{18}$$

The equilibrium decomposition pressures for calcium carbonate are given in Fig. 9.

Although no data have been reported for COS adsorption by half-calcined dolomite, high COS removal efficiencies are predicted thermodynamically according to the reaction:

$$[CaCO_3 \cdot MgO] + COS \rightarrow [CaS \cdot MgO] + 2\,CO_2 \tag{19}$$

where the residual COS concentration is

$$[COS] = [CO_2]^2\ P/K \tag{20}$$

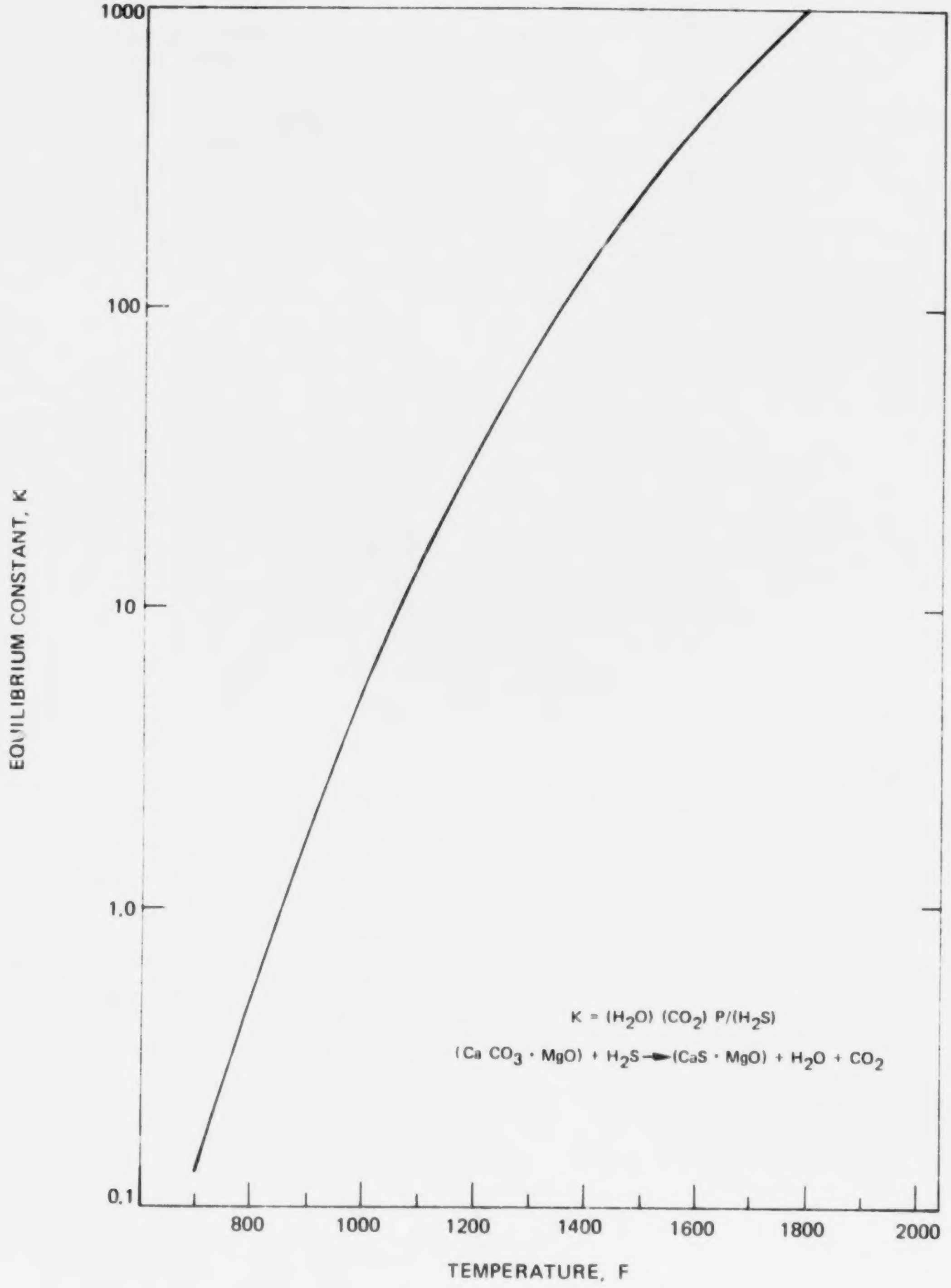
The equilibrium constant for this reaction is shown in Fig. 10.

The process, as described by Consol(20) involves desulfurizing the raw gas in a fluidized bed of half-calcined dolomite acceptor at 1600-1700 F, according to reaction (16). The sulfided acceptor is regenerated by the addition of steam and CO_2 at reduced temperature, thereby reversing the absorption reaction. Regeneration is conducted in a fluidized bed at around 1300 F giving a dilute H_2S off-gas, less than 10% (volume). Because the low H_2S off-gas content prohibits the direct use of a vapor-phase Claus unit for sulfur recovery, Consol is proposing the use of a liquid phase sulfur recovery system based on the Wachenroeder reaction

$$2H_2S + H_2\,SO_3 \rightarrow 3S + 3H_2O \tag{21}$$

FIG. 8

EQUILIBRIUM CONSTANT FOR H_2S ABSORPTION BY HALF – CALCINED DOLOMITE



$K = (H_2O)\ (CO_2)\ P/(H_2S)$

$(CaCO_3 \cdot MgO) + H_2S \rightarrow (CaS \cdot MgO) + H_2O + CO_2$

R1–19–15

FIG. 9

DISSOCIATION PRESSURE FOR CALCIUM CARBONATE

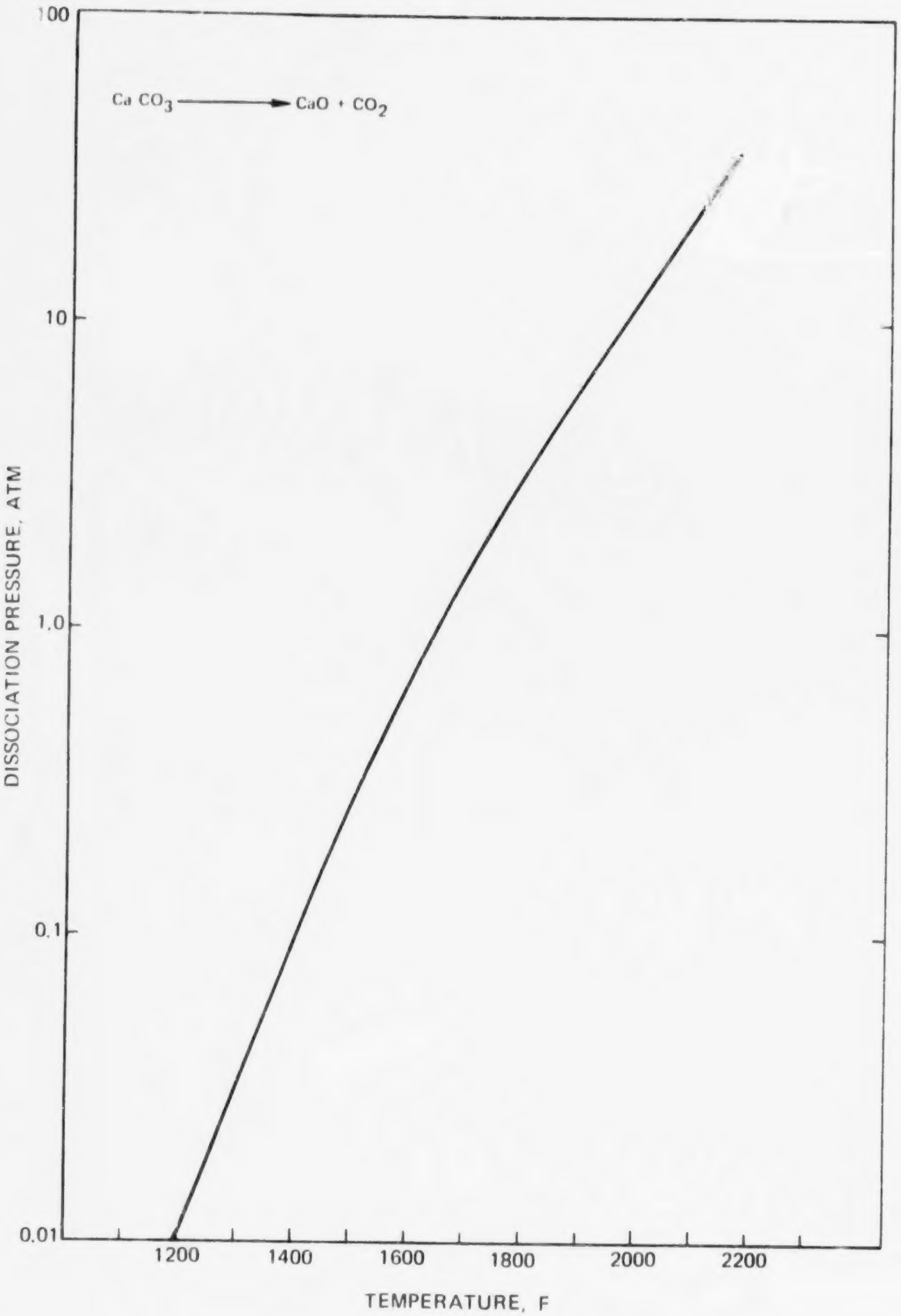


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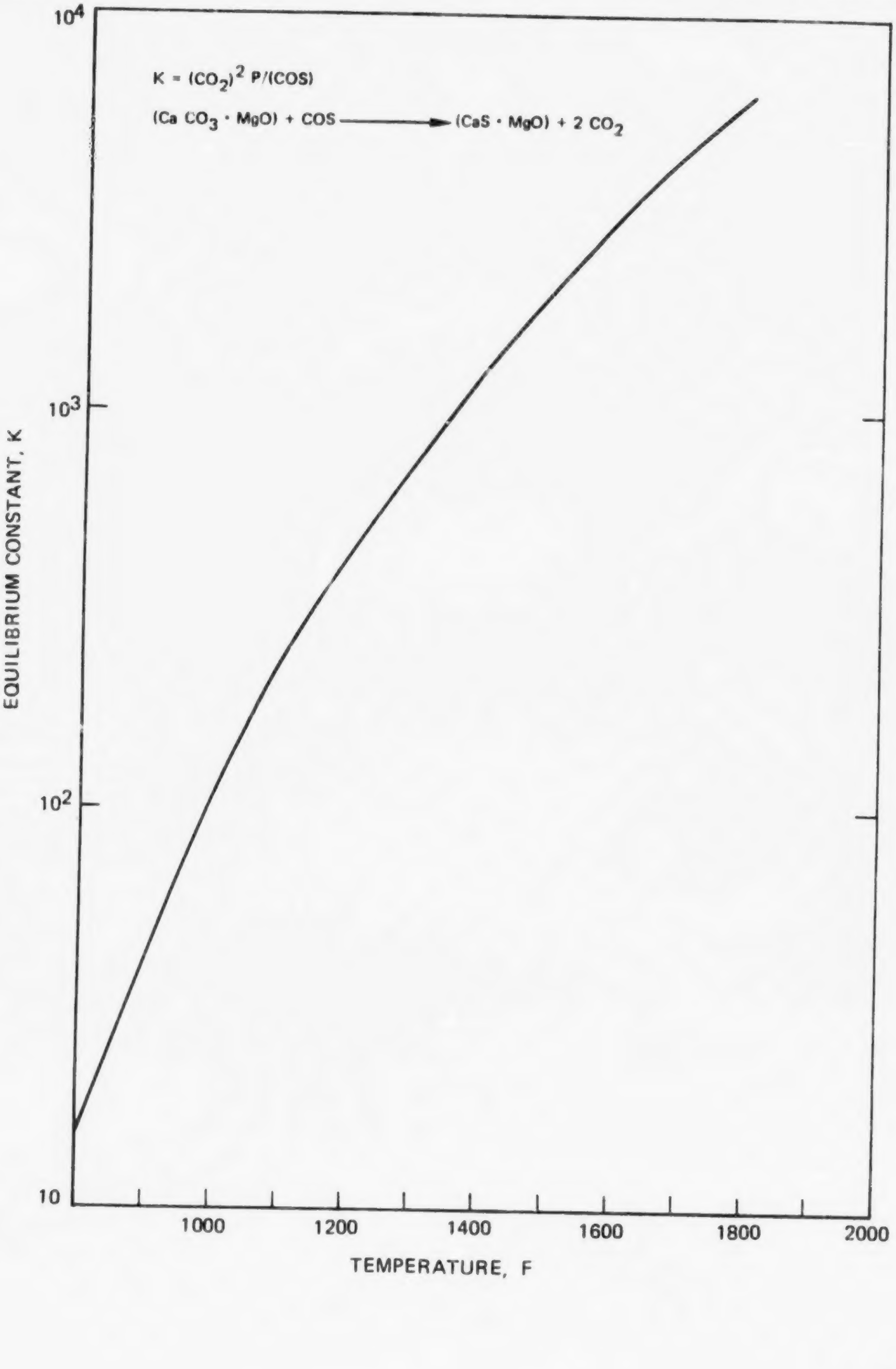
FIG. 10

EQUILIBRIUM CONSTANT FOR COS ABSORPTION BY HALF – CALCINED DOLOMITE



R1–19–16

At the present time, the Consol desulfurization process is under development in the pilot plant stage. An apparent drawback to the process is the low degree of regeneration obtainable for the sulfided acceptor, around 10-13%, which results in a large recirculation of sulfided dolomite to the absorber. For operation at 200 psia, Consol projects a reduction of H_2S from 0.6 percent to 200 ppm in the treated gas.

Air Products Process - As with the Consol process, this system employs dolomite as the sulfur acceptor. However, the acceptor is in the fully calcined form; i.e. $[CaO \cdot MgO]$, and the process therefore consists of three steps: absorption, regeneration, and calcination.

Absorption of hydrogen sulfide takes place at around 1600-1700 F via the reaction with calcined dolomite:

$$[CaO \cdot MgO] + H_2S \rightarrow [CaS \cdot MgO] + H_2O \qquad (22)$$

The equilibrium H_2S concentration may be calculated from the equilibrium constant given in Figure 11 according to the following relationship:

$$[H_2S] = [H_2O] / K \qquad (23)$$

where,

$[\ ]$ = mol fraction of component

K = equilibrium constant

For this system, the H_2S removal efficiency is independent of pressure but, for a given gas composition, it decreases with increasing temperature. From a practical viewpoint, there is a minimum temperature at which this process should operate; that being the temperature at which the decomposition pressure of $CaCO_3$ equals the partial pressure of CO_2 in the gas. Below this temperature carbon dioxide will also be absorbed according to the reaction.

$$Ca\ O + CO_2 \rightarrow Ca\ CO_3 \qquad (24)$$

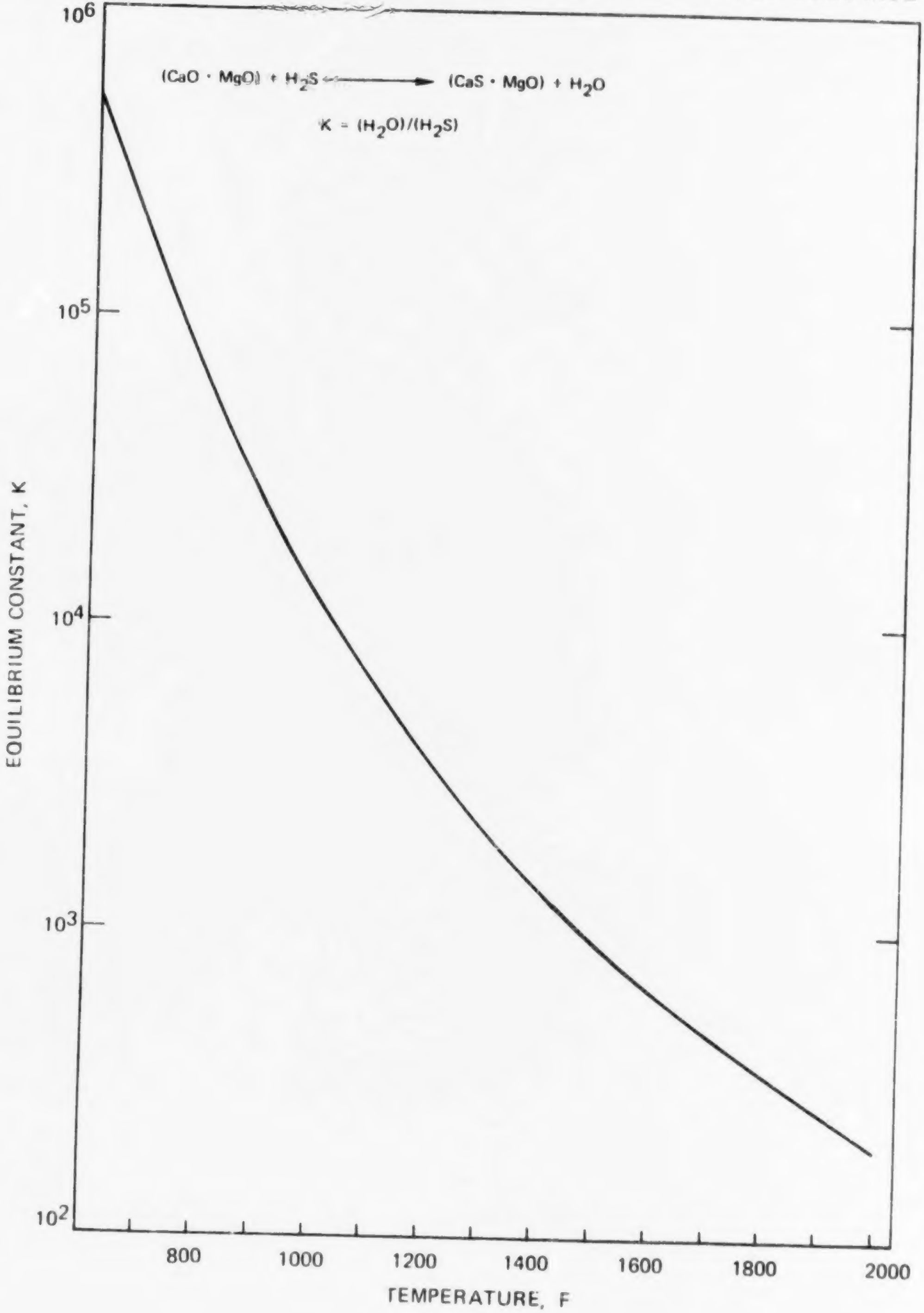
This is detrimental for two reasons:

a. Additional acceptor capacity is required, and

b. the large exothermic heat of CO_2 absorption, 75,000 Btu/mol, will necessitate some means for heat removal from the absorption bed.

FIG. 11

EQUILIBRIUM CONSTANT FOR H_2S ABSORPTION BY CALCINED DOLOMITE



R1-19-22

Like the Bureau of Mines and Consol processes, COS removal appears to be thermodynamically attractive. Residual COS content may be estimated from the chemical equilibrium associated with the absorption reaction:

$$[CaO \cdot MgO] + COS \rightarrow [CaS \cdot MgO] + CO_2 \qquad (25)$$

$$[COS] = [CO_2] / K \qquad (26)$$

where equilibrium constant is given in Figure 12.

Regeneration of the sulfided acceptor is conducted similarly to the Consol process. Steam and carbon dioxide are reacted with the sulfided dolomite at 1100-1200 F resulting in a dilute H_2S off-gas and the carbonated dolomite form.

$$[CaS \cdot MgO] + H_2O + CO_2 \rightarrow [CaCO_3 \cdot MgO] + H_2S \qquad (27)$$

Hydrogen sulfide in the regenerated off-gas may be converted to elemental sulfur via the liquid-phase Wachenroeder process or first concentrated and then fed to vapor-phase Claus unit.

The regenerated dolomite must be calcined before recycle to the subsequent absorption cycle. Calcination is effected at 1900 F with air to drive off carbon dioxide:

$$[Ca\,CO_3 \cdot MgO] \rightarrow [CaO \cdot MgO] + CO_2 \qquad (28)$$

This reaction is endothermic and requires the use of fuel to preheat the air. Excessive temperature, above 2000 F, during calcination can result in deactivation of the acceptor.

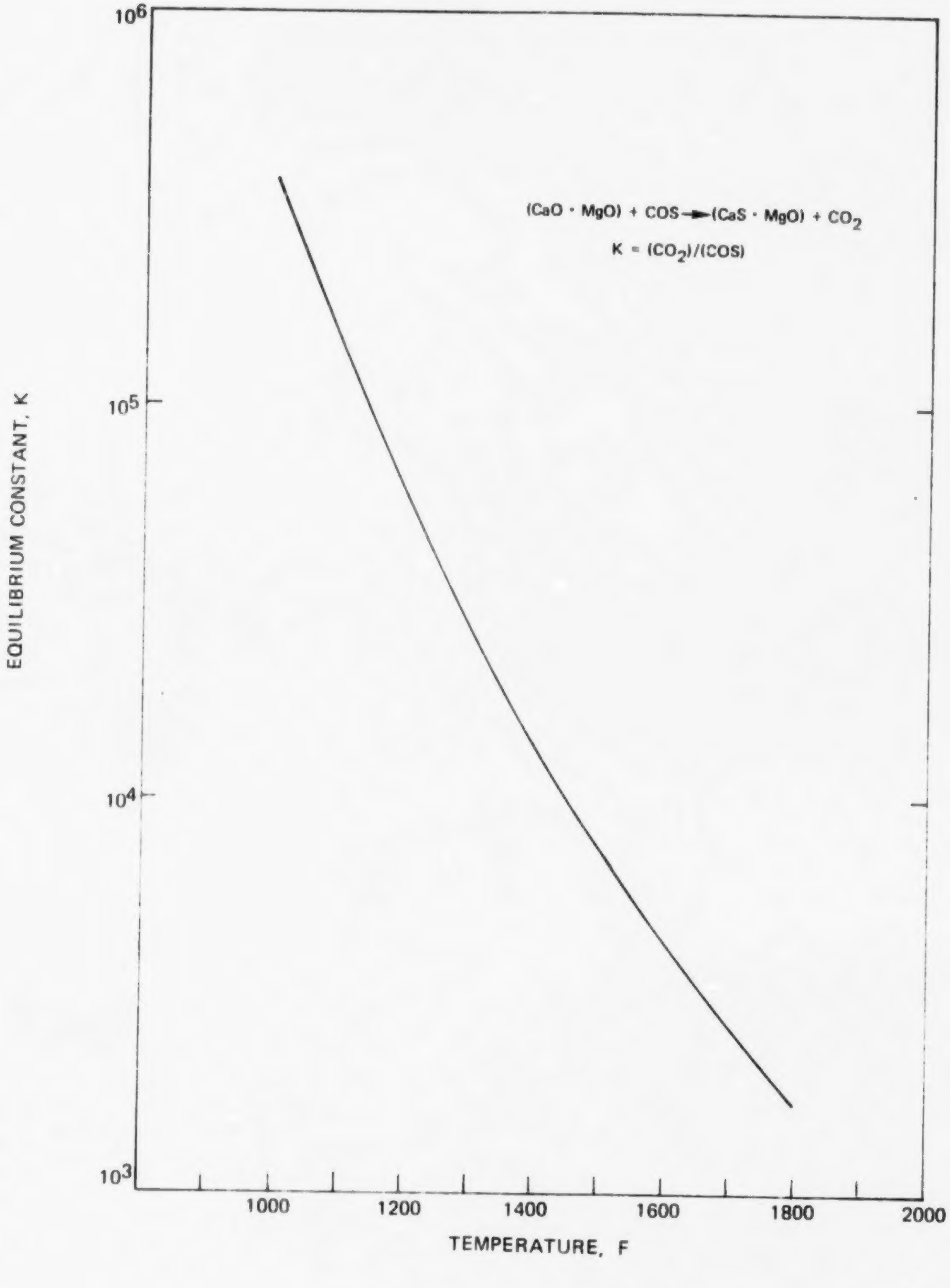
The development of this process by Air Products is currently in the pilot plant stage.(26) The absorption cycle has been successfully tested on actual producer gas. Operating conditions for regeneration and calcination are still under investigation to determine the maximum acceptor life cycles.

IGT-Meissner Process - This process is under development by the Institute of Gas Technology in association with its "U-Gas" process.(27) Details of this technology have not been divulged except that the process operates at 800 F and selectively removes H_2S and COS relative to CO_2.

Battelle Northwest Process - The Battelle process(28) utilizes calcium carbonate, $CaCO_3$, dissolved in a tertiary mixture of alkali metal carbonates to remove H_2S at high temperature. The tertiary carbonate system consisting of potassium carbonate, lithium carbonate, and sodium carbonate, has a eutectic melting point

FIG. 12

EQUILIBRIUM CONSTANT FOR COS ABSORPTION BY CALCINED DOLOMITE



R1-19-11

around 750 F and under operating conditions contains about 15 mol% $CaCO_3$. Besides removing sulfur compounds from the gas stream, this solvent will also scrub out the fly ash constituents from the raw gas.

The system under study at Battelle contacts the molten salt and raw gas in a co-current flow venturi scrubber at temperatures from 1100 to 1700 F. Hydrogen sulfide is removed from the gas by chemical reaction with the carbonate solvent:

$$\mathrm{Ca\,CO_3 + H_2S \rightarrow CaS + CO_2 + H_2O} \qquad (29)$$

The equilibrium H_2S content in the treated gas may be expressed as:

$$[\mathrm{H_2S}] = \frac{(\mathrm{CaS})}{(\mathrm{Ca\,CO_3})} \times \frac{[\mathrm{H_2O}]\,[\mathrm{CO_2}]}{K} \times P \qquad (30)$$

where,

( ) = activity of component in solution

[ ] = mol fraction of component in gas

P = total pressure in atm

K = equilibrium constant

Unfortunately, due to the chemical complexity of the molten salt system, the equilibrium constant cannot be accurately predicted. Observed K values have been a factor of ten below the calculated values. Qualitatively, the H_2S removal efficiency improves with temperature and is inversely proportional to pressure. Experimental data at atmospheric pressure have indicated high H_2S removal, > 94%, at salt loadings under 50% of capacity. Regeneration of the salt is conducted with steam and CO_2 at 1000-1100 F giving an off-gas having H_2S concentrations suitable as feed to a Claus unit for recovery of elemental sulfur.

This process is presently in the pilot plant stage where it will be demonstrated on an actual gasifier raw gas at low pressure, 0-10 psig. It is doubtful that this system can achieve high H_2S removal efficiencies at high pressure and its ability to handle other sulfur contaminants is yet to be demonstrated. Materials of construction for a commercial unit will also present a formidable problem.

Selection Considerations

In general, commercialization of high-temperature systems appears to be 5-10 years away. The processes under development are all capable of selectively removing H_2S down to acceptable levels, less than 500 ppm. Areas which need further definition include ability to contend with other sulfur compounds and contaminants present in raw producer gas, absorbent life expectancy, actual performance data to demonstrate long term reliability, and recovery of elemental sulfur from regenerator off-gases. It would appear that the Bureau of Mines process shows promise of being the first to mature. This process has been demonstrated on actual producer gas, involves relatively simple operation without complex solids handling, and has proven cyclic operation without loss in acceptor activity. The problem of sulfur recovery from the regeneration step represents the major area for refinement.

SECTION 2

PARAMETRIC ANALYSIS OF POWER SYSTEMS

Preliminary analysis of the power systems considered for use with a fuel gasification and cleanup system are presented in this section. They fall into two general categories, steam systems and gas turbine systems. It is increasingly apparent that steam cycles have reached a stage of maturity where little or no improvement can be expected in the next decade. This is best demonstrated by the decline in the percentage of new steam plants using supercritical pressures in the 3500 psi range.(29) The marginal operating cost savings have apparently proven insufficient to justify the added capital costs associated with the higher pressures. Therefore, current steam plant performance will not change significantly with present and future gasifier technology.

Gas turbine technology, on the other hand, has not yet matured and further improvements in the performance of simple-cycle machines can be expected. These improvements generally involve increased turbine inlet temperatures which, when coupled with increased pressure ratios, not only result in better simple-cycle performance but enhance the attributes of the combined gas and steam (COGAS) systems by permitting higher steam temperatures. Both simple- and combined-cycle performance characteristics have been examined to mate with first- and second-generation gasifier technology.

As a point of departure, the performance of a coal-fired steam system with stack gas cleanup has been included in this analysis and in the subsequent economic evaluation as a source of relative comparison.

Since one of the objectives of the program is to identify the differences between and advantages of low- and high-temperature cleanup systems, it was decided to fix the fuel input at a given level and allow the power output to vary. In this way, any scale effects for the gasifier/cleanup systems would be negated. This gasifier plant size has been selected to handle a coal input of 700,000 lb/hr, the amount nominally required for a 1000-Mw coal-fired station. The resultant electrical output is a function of system conversion efficiency and was found to vary from about 800 to over 1,100 Mw. In the combined-cycle generating systems, an even number of gas turbines has been assumed of the largest size practicable using a 3600-rpm single-end power turbine. For these systems, steam turbine size is maintained at or greater than 100 Mw. When the gasified fuel is used directly in conjunction with a steam cycle, performance is based on the use of two 500-Mw units. These factors are discussed in more detail in the following paragraphs of this section.

DIRECT COAL-FIRED STEAM CYCLES

Because of the commanding position currently maintained by steam power systems, they are the logical yardstick against which the feasibility of alternate generation systems must be compared. This section first presents estimates of the performance that can be expected from a typical late 1970-decade steam system without any type of emission controls. The effect of sulfur emission control by stack gas cleanup on that basic plant is then estimated, providing a basis against which the various combinations of gasifiers and cleanup systems can be measured when used in conjunction with either a steam cycle or in advanced power cycles.

The reference steam power plant consists of a turbine-generator and steam cycle mated to a coal-fired boiler. Overall performance for that system is presented in Table 7. The steam cycle used is typical of systems currently under construction with 1000-F, 2400-psi throttle conditions and a single reheat to 1000 F. This same steam cycle is used in evaluating gasified coal-fired steam systems. The significant steam cycle parameters are shown in Figure 13. Without environmental restrictions, a coal-fired system can be expected to have a heat rate of about 9235 Btu/kwh based on the higher heating value (HHV) of the fuel. Of course this increases with condenser temperature and as other auxiliary power requirements increase. With stack gas cleanup, the power required for scrubber operation and the necessary reheat of the stack gas would reduce output power by 5% and give a net heat rate of 9721 Btu/kwh. This varies with the amount of sulfur in the fuel and with the removal process employed. Throwaway systems will have a slightly lower power consumption than those using a regenerable sorbent, but are faced with a severe disposal problem. The cleanup penalty assumed herein, is intended to be representative of systems that will be operational in the 1970's.

The system performance assumes that the yearly average condenser pressure is 2.0 in. Hg when rejecting heat to a wet cooling tower. During the months of June through September the average condenser pressure will be about 3.5 in. Hg. This will result in an increase in the net steam cycle heat rate from 7,836 to 8,058 Btu/kwh or 2.83 percent. This same percent increase applied

TABLE 7

CHARACTERISTICS OF BASIC COAL-FIRED STEAM POWER PLANT

500-Mw Nominal Size Unit

Single Unit Characteristics

Steam Conditions

Throttle Pressure	2,400 psia
Throttle Temperature	1,000 F
Reheat Temperature	1,000 F
Condenser Pressure	2.0 in. Hg.
Final Feedwater Temperature	480 F

Steam Cycle

Net Steam Cycle Output Power	525,000 kw
Auxiliary Turbine Driven Feed Pump Power	11,779 kw
Gross Steam Cycle Power	536,779 kw
Gross Heat Rate	7,664 Btu/kwh
Net Steam Cycle Heat Rate	7,836 Btu/kwh

Station Efficiency

Boiler Efficiency (Illinois #6 Coal) (HHV)	88.5 %
Plant Auxiliary Power	21,630 kw
Net Unit Output	503,370 kw
Plant Heat Rate (HHV)	9,235 Btu/kwh
Efficiency (HHV)	36.96 %

Effect of Stack Gas Cleanup

Reduction in Net Plant Output	5 %
Plant Rate (HHV)	9,721 Btu/kwh
Efficiency (HHV)	35.1 %

BASIC STEAM CYCLE FLOW DIAGRAM

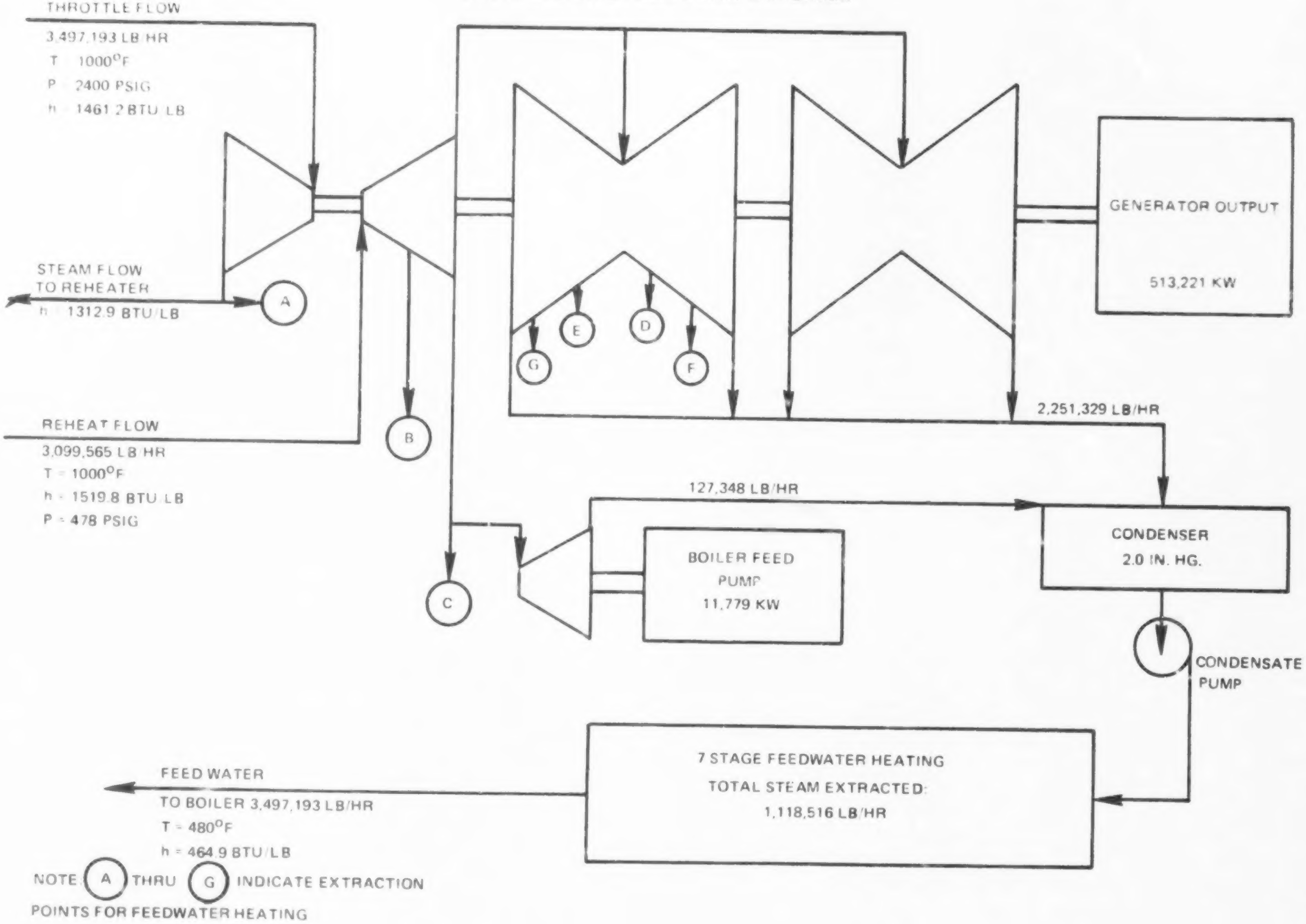

THROTTLE FLOW
3,497,193 LB/HR
T = 1000°F
P = 2400 PSIG
h = 1461.2 BTU/LB
STEAM FLOW
TO REHEATER
h = 1312.9 BTU/LB
A
REHEAT FLOW
3,099,565 LB/HR
T = 1000°F
h = 1519.8 BTU/LB
P = 478 PSIG
B
C
G
E
D
F
GENERATOR OUTPUT
513,221 KW
2,251,329 LB/HR
127,348 LB/HR
BOILER FEED
PUMP
11,779 KW
CONDENSER
2.0 IN. HG.
CONDENSATE
PUMP
7 STAGE FEEDWATER HEATING
TOTAL STEAM EXTRACTED:
1,118,516 LB/HR
FEED WATER
TO BOILER 3,497,193 LB/HR
T = 480°F
h = 464.9 BTU/LB
NOTE Ⓐ THRU Ⓖ INDICATE EXTRACTION
POINTS FOR FEEDWATER HEATING

FIG. 13

RO2-187-1

to the overall plant heat rate during those months, raises it from 9721 to 9996 Btu/kwh.

Heat rates presented here and used throughout the study were based on performance at rated plant output. Typical operation at 70 to 80 percent of capacity would produce a five percent increase in these values for conventional steam plants. No data are available for the corresponding factor in operation of a gasified coal system, but it is assumed that losses associated with less than full output will be similar to the steam station experience.

Design Considerations - The design characteristics of future steam power stations have been examined in some detail.(3) The principal factors affecting steam station performance are unit size, steam conditions, and environmental and fuel factors all of which must be considered in the design.

Unit Size - The overall size of the electric utility system generally fixes the size of new units or at least limits them to about 10 percent of the total system size. Therefore, a number of large utilities could be expected to install units of over 1000 Mw in size. However, a survey (29) of about 50 percent of the new plants scheduled for completion in the 1975 to 1978 time period shows no single unit having a capacity of over 1000 Mw and shows an average capacity of 494 Mw for conventional steam units. Because of this, the basic steam plant has been assumed to consist of a pair of 500 Mw units.

Steam Conditions - The general trend in steam conditions has been toward higher pressure and temperatures to produce power more efficiently. When combined with reheating and regenerative feedwater heating, significant efficiency improvements in the basic Rankine cycle have been made possible. However, material capabilities, additional costs, and diminishing returns on both initial steam condition increases and number of reheats place a practical economic limit on the degree to which these changes can be used to improve performance. A 1000 F peak steam temperature is generally accepted as the most practical level for the foreseeable future. The increased material costs and poor operating experience resulting from higher temperatures are primarily responsible for the limitation. Virtually all of the plants surveyed(29) have steam temperatures at or near 1000 F. Steam pressure, on the other hand, has not found a universally accepted level. However, there is a trend downward toward the 2400-2600 psi level from the 3500 psi supercritical level that was used in about 50 percent of new plants in the early 1970's. Only 24 percent of the conventional plants due to start operation through 1978 will use supercritical pressures.

The use of a second reheat has also declined in new systems with none planned for the next four years. This is understandably tied to the decline in use of supercritical pressure systems as well as the small increase in performance afforded by the second reheat as shown in Figure 14.

Exhaust Conditions - Performance of steam equipment is highly sensitive to available cooling water temperatures. In spite of this, over 50 percent of currently planned stations will rely on cooling towers and about 40 percent will use cooling ponds or lakes. The minimum practical condensing temperature is about 42 F above the dew point temperature for a wet cooling tower. This differential results from a 10 F terminal difference in the condenser, a 20 F drop across the tower and a minimum approach differential of 12 F. These factors are discussed by Burns and Roe(30) who also identify a wet bulb temperature of 75 F as the average during the months of June through September with a peak temperature of 83 F. This results in an average condenser pressure of 3.5 in. Hg during that period and a peak (exceeded only 1 percent of the time) of 4 in. Hg.

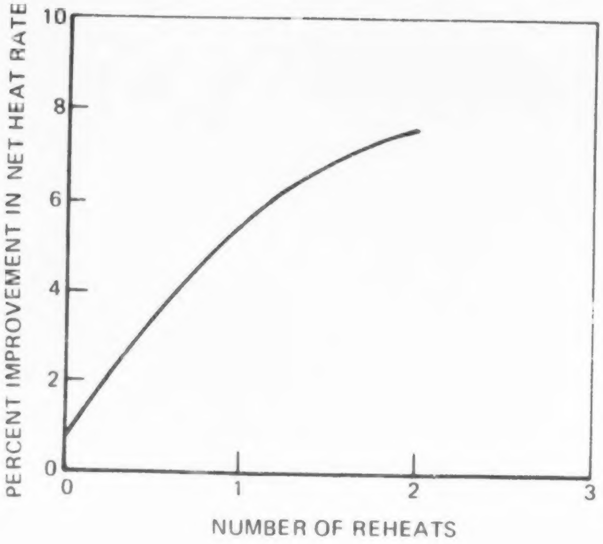
The use of cooling towers has undoubtedly had an effect on the design of new units. The sensitivity of heat rate to changes in condenser pressure can vary widely with plant design. The basic system selected as a standard shows a moderate sensitivity, its heat rate increasing by 2.8 percent as condenser pressure is increased from 2.0 to 3.5 in. Hg. This sensitivity is a function of the ratio of steam flow to last-stage annular area. The higher the ratio the lower the sensitivity. However, with a high ratio of flow to annular area, it is not possible to take advantage of low condenser pressures when they are available. Thus, it would appear that the use of cooling towers will cause a trend toward slightly higher flow to area ratios with a lower cost per unit output and slightly lower efficiency.

Environmental Considerations - Heat rejection and stack gas cleanup for SO_2 have been the primary areas of concern in preventing the degradation of the environment. The first of these is highly dependent on the plant location so that the method of heat rejection will vary widely. Performance characteristics assume the use of a mechanical draft cooling tower.

Performance of stack gas cleanup systems are much more difficult to characterize because of the widely varying results of the various test systems. Therefore, this study did not attempt to select a particular design nor was the topic reviewed in great detail. The net power losses attributable to the combination of cooling towers and stack gas cleanup are given (29) at between 6.4 and 6.7 percent of net capability. Another estimate, (31) places losses due to stack gas cleanup at between two and seven percent depending on the process and design philosophy. A value of five percent was chosen to be

FIG. 14

EFFECT OF REHEATING



R04–12–1

representative and shows good correlation with data for a regenerable scrubbing system. (32)

Auxiliary Power - While not a major factor in plant efficiency or operating cost, the power consumed by auxiliary equipment can be between three and six percent of the net steam cycle output. When comparing different types of plants having different auxiliary power characteristics it is important that these be quantified so that a meaningful comparison of plant efficiency and operating cost can be made. Table 8 presents a typical breakdown of auxiliary power required for a coal-fired steam plant. To permit uniform application of these factors to each of the other systems to be considered, the auxiliaries have been broken into five categories. The portion shown for coal transport and pulverization is used only in the basic steam plant. In the gasified-coal plants a separate estimate of power for coal preparation has been made. This is included in the bookkeeping for the integrated system and, in the case of the BCR gasifier , amounts to about 2.27 percent of the net station output. Boiler fan power has been deducted from both the basic steam plant and the gasified coal-fired steam plant. All the advanced-cycle plants run with a back pressure on the gas turbine sufficient to meet the pressure drop across a waste heat boiler. Condensate pump power has been deducted from all plants based on the net steam power developed. The power for cooling water pumps has also been applied to the steam power, but in the gasified coal systems an adjustment has been made to account for the cooling water requirements of the cleanup system. The miscellaneous plant losses include lighting and ventilating as well as copper (electrical resistivity) losses and apply equally regardless of the primary power cycle.

GASIFIED COAL-FIRED STEAM SYSTEMS

A potential alternative to stack gas cleanup is to clean the fuel prior to firing. This can be done by first gasifying the fuel and then cleaning it in one of the many systems described in other sections of the report. The primary advantage of such a system over stack gas cleanup is the lower quantity of gas (only about 40 percent of the stack gas quantity) that must be processed. Also, if the cleanup system is run at pressure, gas volume is further reduced and the sulfur compounds can be removed more readily by a wide variety of processes. Finally, sulfur leaving the gasifier is mostly in the form of H_2S which, after separation, can be more easily converted to elemental sulfur than can SO_2.

Table 8

AUXILIARY POWER REQUIREMENTS

Percent of Net Steam Cycle Output

*Coal Pulverizing and Transport	.81%
Boiler Fans	1.51%
Condensate Pumps	.22%
Cooling Water Pumps	1.01%
Miscellaneous Plant Losses	.57%

*Applicable only to direct coal fired boiler. Estimated separately for gasified coal systems based on gasifier requirements.

System Configuration

The general system arrangements that would be used in a gasified coal-fired steam system are shown in Figures 15(a) and 15(b). With a low-pressure gasifier and cleanup system (Figure 15(a)) the remainder of the plant consists of a gas-fired boiler generating steam for use in a conventional steam cycle as was shown in Figure 13. The low fuel heating value will require the burners to handle a volumetric flow of fuel some six times that of a natural gas-fired boiler. While conventional boiler design practices are applicable to a low-Btu system, lower flame luminosity and higher stack gas flows as compared to coal and natural gas fired boilers will result in a different split between radiation and convection heat transfer surface required. As a result, use of low-Btu gas in existing boilers would require some derating and modification to maintain good control of steam conditions. However, new designs would display little outward difference from existing boilers. For pressurized operation of the gasifier and cleanup system it is necessary to compress the gasifier inlet air as shown in Figure 15(b). After cleanup, the resultant high-pressure gas can be expanded through a turbine to recover some or all of the work of compression. If the temperature of the gas entering the turbine is high enough, power can be extracted from the let-down turbine. Depending on the ratio of product gas to air flow into the gasifier, a fuel temperature in the range of 600 to 800 F will be sufficient to sustain the compressor and a greater temperature will produce net output power from the let-down turbine.

System Performance - For comparison, the preliminary estimates of the performance of the steam systems are given in Table 9.* Two gasifier/cleanup combinations were considered; a Bu Mines/low-temperature system and a BCR/high-temperature system. Steam cycle performance, similar to that in Table 7 was used for all system combinations. Steam boilers firing gasified fuel have lower efficiency than direct-fired boilers because of the steam injected into the gasifier during the gasification process. Some of the steam reacts with coal to produce H_2 and CH_4 which form H_2O upon combustion and results in higher moisture losses. That steam which does not react with coal and is not removed from the fuel gas passes through the boiler to the stack and causes a slight increase in the stack gas sensible heat loss. The boiler efficiency estimates presented in Table 9 were calculated using standard boiler practice.(33)

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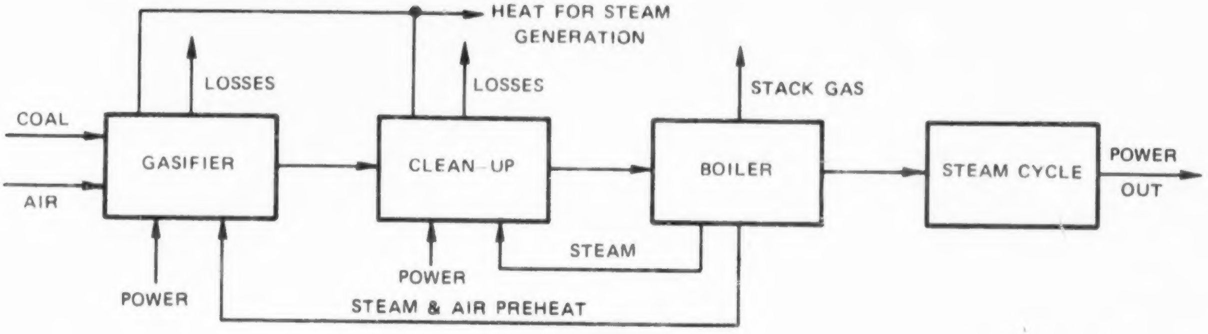
* Numbers in Table 9 are early estimates and are not entirely consistent with numbers in Table 7.

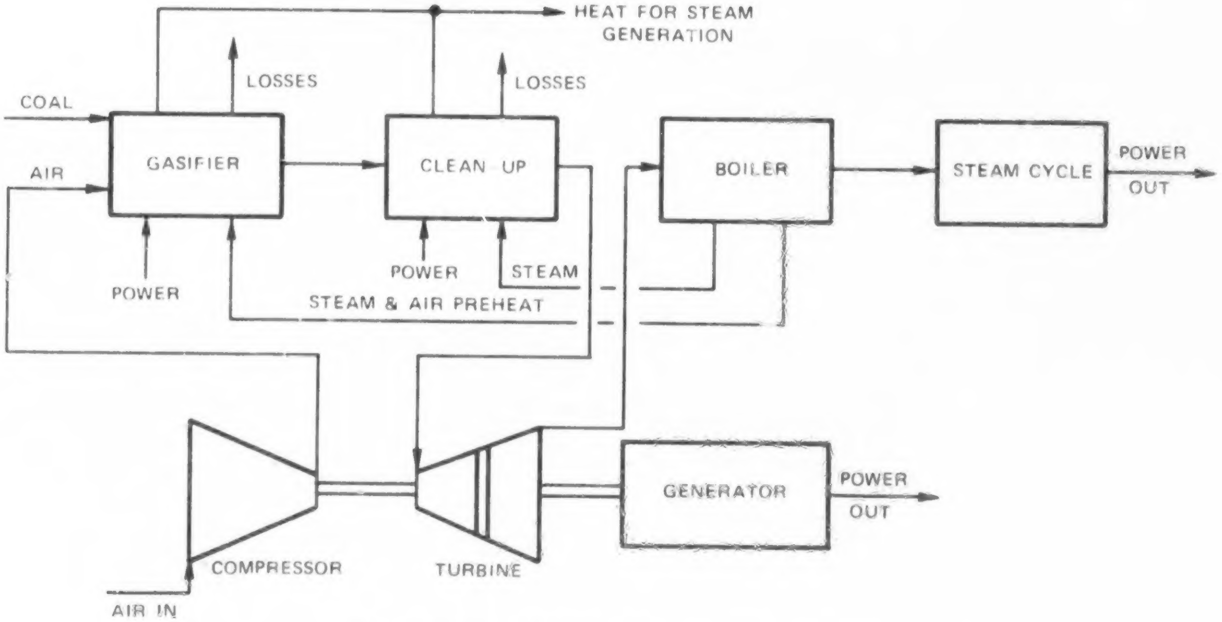
FIG. 15a

GASIFIED COAL–STEAM CYCLE SYSTEMS


HEAT FOR STEAM GENERATION
LOSSES
LOSSES
STACK GAS
COAL
AIR
GASIFIER
CLEAN–UP
BOILER
STEAM CYCLE
POWER OUT
STEAM
POWER
POWER
STEAM & AIR PREHEAT

LOW PRESSURE GASIFIER

FIG. 15b


HEAT FOR STEAM GENERATION
LOSSES
LOSSES
COAL
AIR
GASIFIER
CLEAN UP
BOILER
STEAM CYCLE
POWER OUT
POWER
POWER
STEAM
STEAM & AIR PREHEAT
GENERATOR
POWER OUT
COMPRESSOR
TURBINE
AIR IN

HIGH PRESSURE GASIFIER WITH LET DOWN TURBINE

TABLE 9

STEAM POWER STATION EFFICIENCY COMPARISON

		Steam Cycle	Boiler(1)	Gasifier	Station(2)
I.	W. Kentucky No. 9 Coal Direct Firing	43.5	86.9	---	35.5
	Direct Firing/SGC	43.5	82.55(3)	---	33.8
	BuMines Gasifier/ Selexol Cleanup	43.5	84.8	77.0	26.7
	BuMines Gasifier/ Cleanup @ 1000 F(4)	43.5	85.2	84.1(5)	29.3
II.	Illinois No. 6 Coal Direct Firing	43.5	88.5	---	36.2
	Direct Firing/SGC	43.5	84.1(3)	---	34.4
	BCR Gasifier/ Selexol Cleanup	43.5	84.3	88.0	20.3
	BCR Gasifier/Cleanup @ 1000 F	43.5	85.5	89.4(5)	31.3
	BCR Gasifier/Cleanup @ 1800 F(4)	43.5	86.8	87.8(5)	31.2
	BCR Gasifier/Turbo-Compressor Cleanup @ 1500 F	43.5	85.3	88.3	31.8

(1) 300 F stack, 59 F ambient, 20% excess air for direct firing, 10% excess air for gasified fuel firing.
(2) 6% auxiliary power loss
(3) Includes 5% loss for SGC system
(4) Gasifier exit temperature
(5) Neglects fuel and steam requirements for cleanup and regeneration systems

The principal gasification/cleanup process losses are due to unconverted carbon from the coal and the need to quench the fuel gas to remove tars and/or to cool the gas prior to desulfurization. For the purpose of making steam system performance comparisons, an effective gasifier efficiency was used. This efficiency is defined as the net energy delivered to the steam boiler divided by the gross input energy in the coal. The net energy delivered to the boiler includes fuel gas sensible and chemical energy, plus an adjustment for heat added directly to (or removed from) the steam cycle. The hot cleanup system assumed is a fictitious one with no parasitic losses (fuel, power, or steam) and which removes only H_2S, COS, and NH_3. Direct comparison of the efficiencies for the two types of gasifiers would be misleading because of the different types of coals used at the different process operating conditions and assumptions.

The performance estimates presented in Table 9 indicate that the efficiency of a steam station using a Bu Mines gasifier and low temperature cleanup would be 8.8 points below the efficiency of a direct fired station. This decrement would be reduced to 6.2 points if a 1000-F cleanup system could be used. Similarly, the efficiency decrement for a BCR gasifier with low-temperature cleanup would be 5.9 points and this would be reduced to 4.9-5.0 points for cleanup at 1000-1800 F. The addition of stack gas cleanup reduces the differential by 1.7-1.8 points.

In the preceding estimates it was assumed that the power required to drive the booster compressor which feeds the gasifier could be supplied by an expander turbine with matching output power. An equivalent assumption would be that the gasifier is operated at low pressure so that the compressor and turbine powers would be negligible. The validity of this assumption was tested for the BCR gasifier operating at 500 psi. Fuel gas from the gasifier was cooled to 1500 F, desulfurized, and expanded through the turbine to 20 psi before being burned in the boiler. The 1500 F turbine temperature limit was imposed because of the difficulty of cooling a fuel expander turbine. (Cooling air could not be used because combustion would occur in the turbine.) In this case the expander turbine was capable of delivering net power amounting to approximately 3 percent of the net station output. The net station efficiency in this case, was .5-.6 points higher than for the previous case which did not incorporate a turbo-compressor.

It appears from this analysis that the benefits to be derived by using hot cleanup in conjunction with a BCR gasifier and a steam power cycle are marginal in comparison to low-temperature cleanup, whereas the benefit for a Bu Mines-type gasifier would be significant. This is because the heat recovery in the BCR/low temperature went into steam at essentially cycle efficiency just as does the sensible heat in BCR/high-temperature. The

Bu Mines/low-temperature system has essentially no heat recovery. In any event, it would be extremely difficult to envision such a system with a net efficiency much over 30 percent. This result is consistent with an estimate of 27 percent presented by Commonwealth Edison Company.(34)

ADVANCED COGAS SYSTEMS

Several studies(34) of advanced combined gas and steam (COGAS) cycles conducted by the United Technologies Research Center has shown that these cycles offer the potential of high performance with low capital investment. They are particularly well adapted to use with gasified fuels where gasifier and cleanup systems operate at pressure. A simplified schematic of a COGAS system is shown in Figure 16.

In the above arrangement, the turbine exhaust gas is used directly to raise steam in an unfired boiler. This arrangement has been shown (3) to have the potential for highest efficiency and lowest overall cost. For a given stack temperature, T_{st}, the efficiency of a COGAS system is a function of four primary variables: the gas turbine efficiency, η_{gt}; the steam cycle efficiency, η_s; the turbine exhaust temperature, T_{ex}; and supplemental fuel, W_s. The relationship is:

$$\eta_{cc} = \frac{\eta_{gt} + [(1 - \eta_{gt}) + \frac{W_s}{W_f}](\frac{T_{ex} - T_{st}}{T_{ex} - T_a})\,\eta_s}{1 + \frac{W_s}{W_f}} \qquad (31)$$

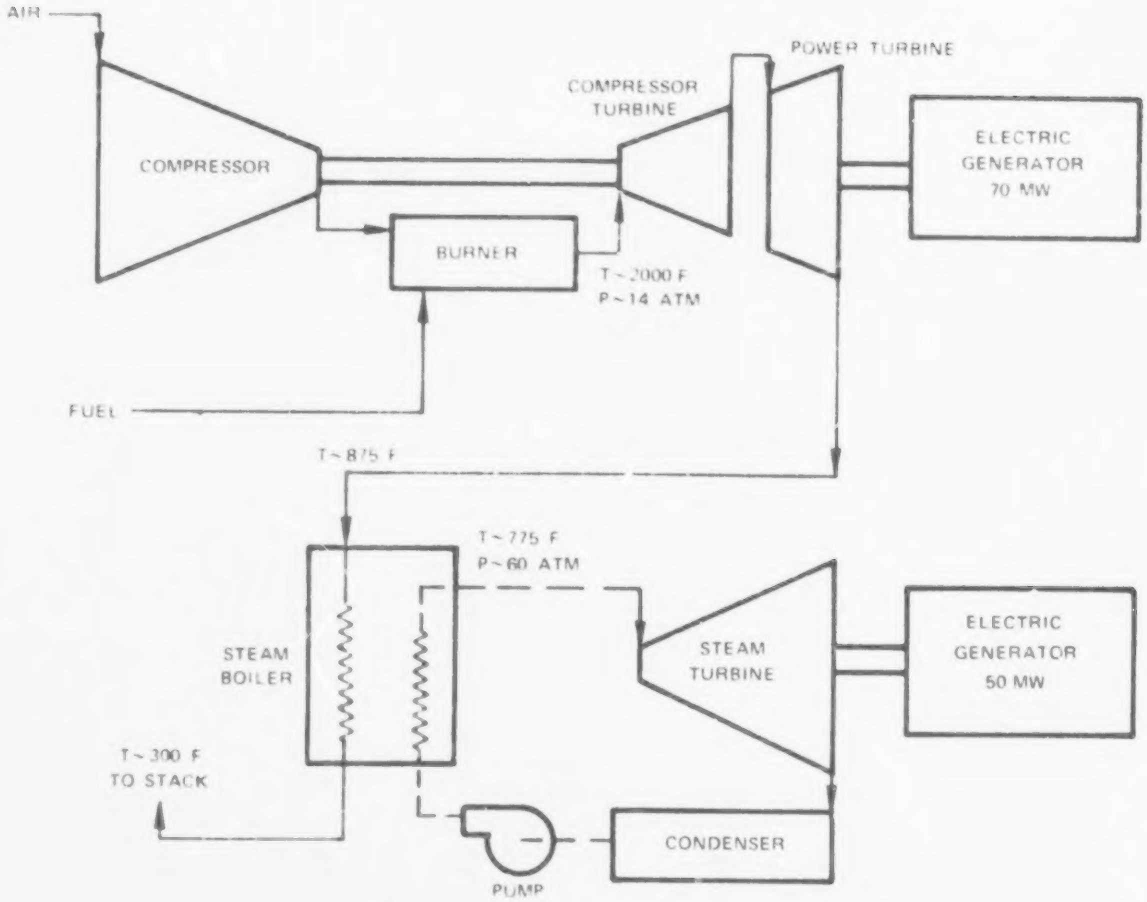
Other terms in Equation 31 are engine fuel flow rate, W_f; and the ambient temperature, T_a. The performance of a simple waste-heat-recovery system with no supplementary firing is then:

$$\eta_{cc} = \eta_{gt} + (1-\eta_{gt})\,\eta_s\,\frac{T_{ex} - T_{st}}{T_{ex} - T_a}\,. \qquad (32)$$

This equation points out the importance of the basic gas turbine efficiency and of the cycle temperatures on the ability to utilize the gas turbine exhaust heat. Exhaust temperature has a two-fold effect on the steam cycle contribution to combined cycle efficiency. It not only improves the ratio of available to total heat in the gas stream but higher temperatures allow the use of higher performance steam cycles. For these reasons, it has been relatively common practice in currently installed combined-cycle systems to use supplementary fuel to increase boiler inlet temperature and improve overall cycle efficiency.

FIG. 16

WASTE–HEAT COMBINED GAS AND STEAM TURBINE SYSTEM


AIR
COMPRESSOR
COMPRESSOR TURBINE
POWER TURBINE
ELECTRIC GENERATOR 70 MW
BURNER
T ~ 2000 F
P ~ 14 ATM
FUEL
T ~ 875 F
T ~ 775 F
P ~ 60 ATM
STEAM BOILER
STEAM TURBINE
ELECTRIC GENERATOR 50 MW
T ~ 300 F
TO STACK
PUMP
CONDENSER

Similar changes in steam cycle contribution to performance could result from changing either the gas turbine pressure ratio or turbine inlet temperature. It is the objective of this section to illustrate the interdependency of the gas turbine and steam cycles and to provide a framework for the integration of the gasifier and cleanup systems into an advanced power station.

Gas Turbine Cycle

Considering overall COGAS system performance, the single parameter that can improve both gas turbine and steam cycle performance is gas turbine inlet temperature. In the following paragraphs the various approaches to improving this important parameter are described.

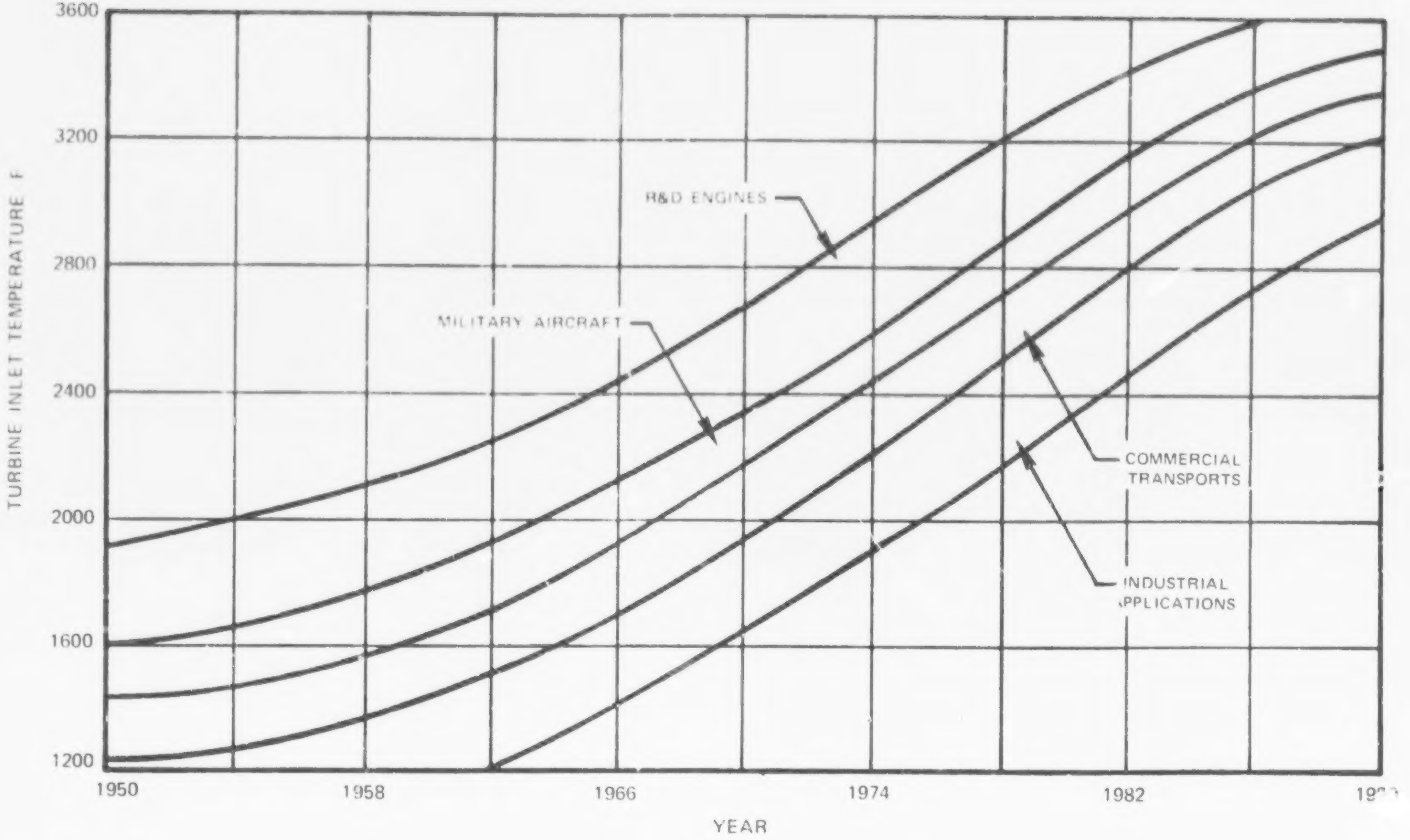
Turbine Inlet Temperature - The current state of the art in gas turbine design has seen the aerodynamic efficiencies of the turbine and compressor reach 90 percent or better. While higher aerodynamic efficiencies are desirable, the single most effective method of increasing the overall performance of the turbine is to increase the turbine inlet temperature (TIT). The historical trend in TIT's is depicted in Figure 17. After high-temperature technology is tested and proven in research and development engines, it is first applied to military aircraft engines, then to commercial aircraft engines, and finally, some 6 to 8 years after the R&D engine, to industrial engines. A close look at Figure 17 will also show that the technology transfer time, i.e., from R&D engine to industrial engine, is becoming shorter. Thus, the advances in cooling techniques and materials first used on research engines will appear much sooner in industrial engines than they have in the past. It may be, in fact, that certain cooling techniques, e.g., water-cooled blades, would bypass the aircraft application altogether.

A number of blade cooling techniques and improved materials could be considered for use in future power systems. For this study, five combinations of materials and cooling techniques were selected as representative of future technology and their performance was studied in detail for turbine inlet temperatures of 2200, 2400, 2600, and 3000 F. The five combinations are:

I. Conventional air cooling
II. Conventional materials but vane cooling air precooled to 300 F
III. Ceramic vanes and conventionally cooled blades
IV. Ceramic vanes and blades
V. Zero bleeds and leakage

FIG. 17

ESTIMATED TURBINE INLET TEMPERATURE PROGRESSION



Cooling Case I corresponds to using conventional cooling techniques and conventional materials maintained at a maximum temperature of 1500 F for long lifetime, baseload use. In a typical gas turbine, a number of bleed streams of air are extracted from the compressor for cooling purposes in the hot turbine sections. Most of the cooling air is used in the first pair of vanes and blades (see Figure 18). This cooling air is at a temperature corresponding to the compressor exit and is a function of cycle pressure ratio. Cooling air flow for the second pair of vanes and blades is on the order of 1/3 of the flow to the first pair. The remaining bleed flows for disc, bearing, and miscellaneous cooling typically amounts to 6 percent. For Cooling Case II the cooling flow to the vanes is independent of cycle pressure ratio and varies, for the first vanes, from 7.8 percent at 2200 F to 11.0 percent at 2900 F. Case III incorporates ceramic vanes (no vane cooling), but the remaining cooling flows are the same as for Case I. Case IV incorporates ceramic vanes and blades (no vane or blade cooling), but the required cooling flows to the discs, bearings, etc. remain at 6 percent. Case V corresponds to the theoretical limit with no cooling bleeds of any kind.

For first-generation systems, the use of conventional impingement type blade cooling at a TIT of 2200 F. is judged to be representative of engines to be used in the late 1970's. For second-generation systems, the use of ceramic vanes with conventional blade cooling and a turbine inlet temperature of 2600 F was selected. The effect of these combinations on overall combined-cycle system performance (distillate oil-fired systems are used for comparison) is summarized in Figure 19 and is discussed in detail in a later section on COGAS performance. The effects of temperature and cooling configuration on simple cycle performance are shown in Figures 20 and 21. In these figures the gas turbine efficiency is plotted against net power per unit airflow or specific power. This is a measure of the amount of work which can be done by a given size machine and is, indirectly, a guide to cost. A machine with a high specific power will generally have a lower cost in $/kW than a machine of lower specific power. It is interesting to note in Figure 20 that with conventional cooling the incremental improvements in efficiency resulting from increases in turbine inlet temperature are small. The major effect is an increase in specific power. This same general trend is apparent in Figure 21, however, the reduced cooling flows associated with ceramic materials result in a more significant efficiency improvement. It is only when the gas turbine is used in a COGAS system that the full benefit in improved cycle efficiency as well as specific power associated with higher turbine inlet temperatures can be realized.

EFFECT OF CYCLE PARAMETERS ON TURBINE COOLING AIRFLOW REQUIREMENTS

CONVENTIONAL AIR COOLING TECHNOLOGY

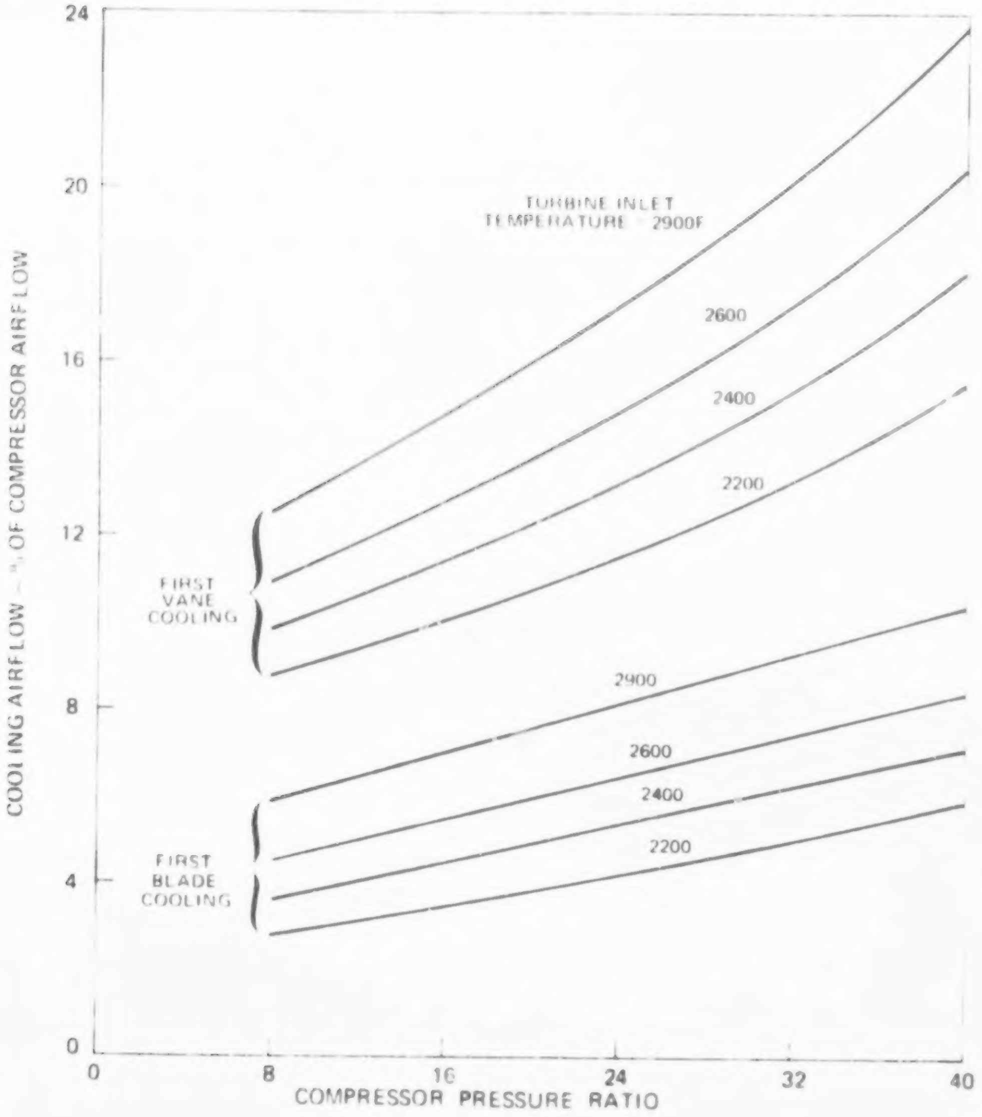


FIG. 19

COGAS STATION PERFORMANCE

DISTILLATE FUEL

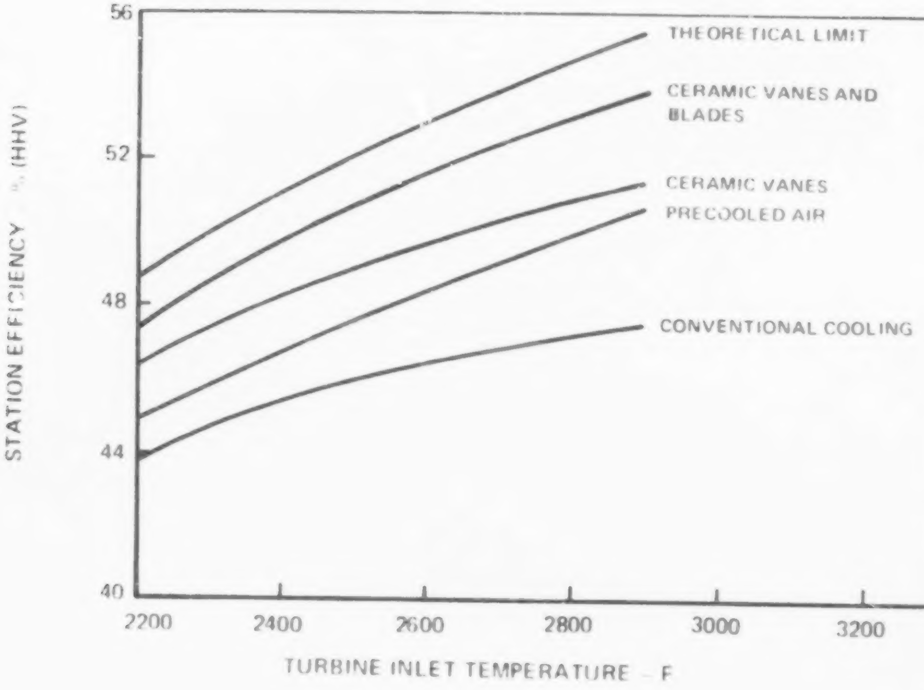


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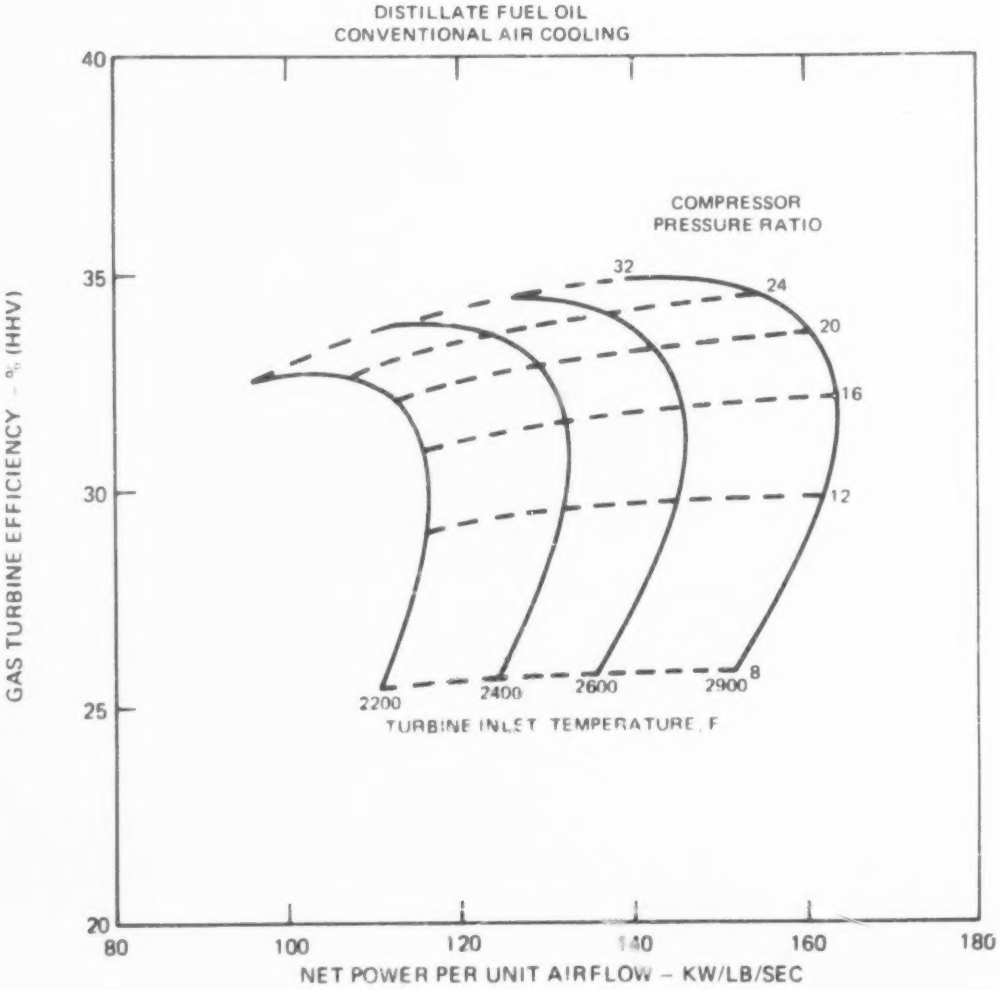
FIG. 20

GAS TURBINE PERFORMANCE MAP



PERFORMANCE OF 2600 F GAS TURBINE

DISTILLATE FUEL OIL

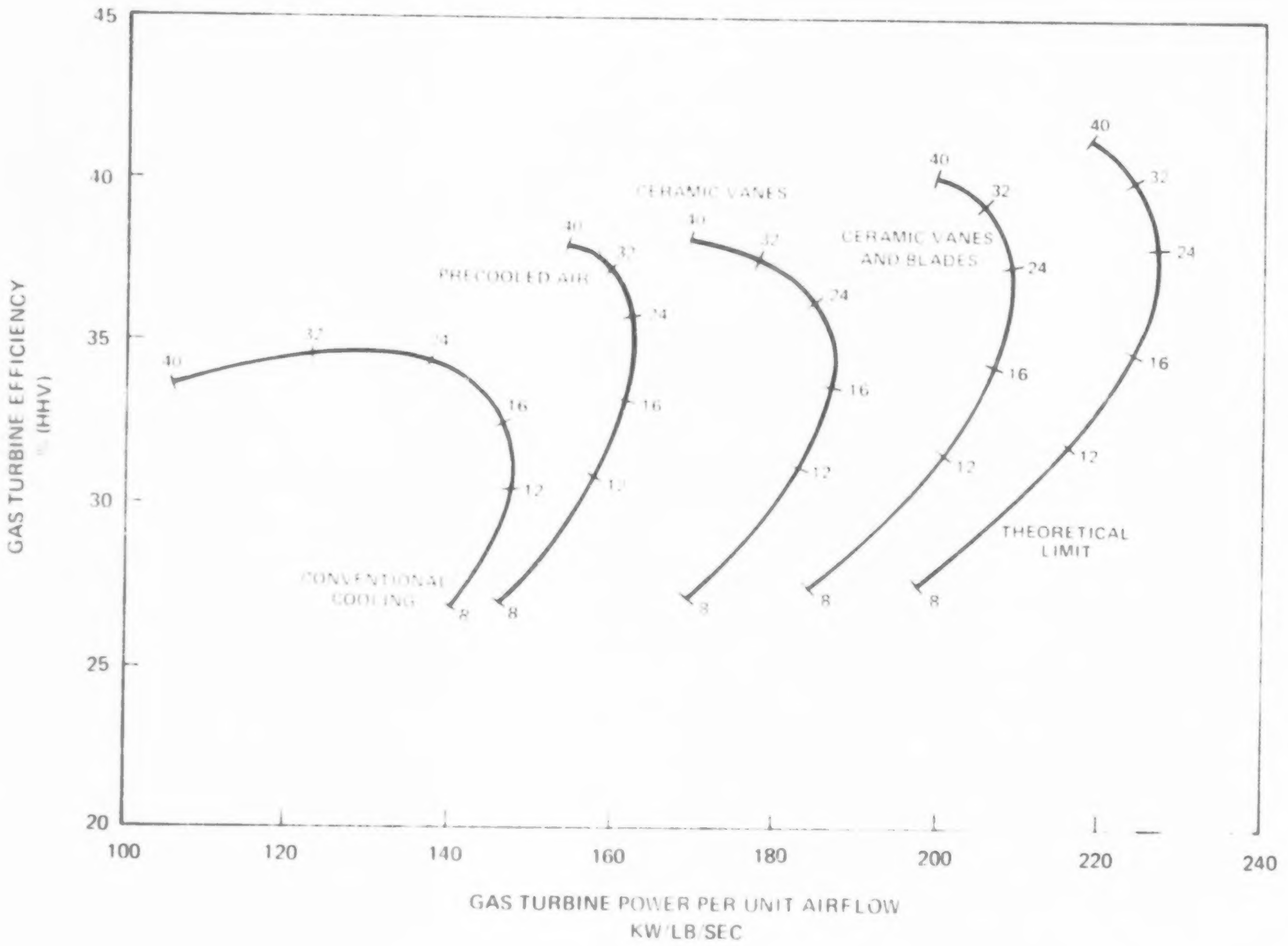


FIG. 21

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Steam System

Although the efficiency term for the steam-cycle in equation (31) is modified by several coefficients, it is apparent that it is desirable to achieve as high an efficiency as practicable. There are, however, constraints on the steam cycle conditions which are external to the steam power system. These are gas temperature to the boiler (gas turbine exit temperature) and the stack temperature. The latter is set by stack gas dew point and is a function of corrosion and other practical considerations. It is usually not much below 300 F, even when a clean fuel is used. While a lower stack temperature could be considered, it represents a study that must be done for each specific system. For this study a 300 F stack temperature was chosen as a standard and all systems compared on that basis. It has been found$^{(4)}$ that a pinch temperature difference between steam and exhaust gases of 100 F results in the most acceptable heat exchanger design. Thus, for systems utilizing waste heat recovery boilers, the maximum steam temperature is 100 F less than the gas turbine exhaust temperature. At the low-temperature end of the waste-heat-recovery unit, a pinch of 50 F is acceptable in the economizer between exhaust gas and feedwater. Thus, to maintain a 300 F stack temperature and utilize all of the available heat in the turbine exhaust, regenerative feedwater heating is limited to a maximum of 250 F.

High-performance steam cycles such as a 2400 psi 1000 F/1000 F cycle would generally require exhaust firing to raise the gas temperature to a level sufficient ([illegible]-1400 F) to accommodate both the initial superheat as well as the reheat section. Without regenerative feedwater heating, such a cycle would have an efficiency of about 39 percent. As the gas temperature into the waste-heat boiler is reduced, steam temperature must be decreased and cycle performance goes down. For example, a 1250 psi 900 F cycle without reheat would have an efficiency of about [illegible] percent. However, even at this reduced steam temperature it is necessary to have a gas turbine exhaust temperature of 1100 to 1200 F to produce a stack temperature of 300 F. Lower exhaust temperatures result in higher stack temperatures for a fixed steam cycle. This effect is shown in Figure [illegible]. When using a single-pressure steam system, the gas side temperature profile pivots about the evaporator pinch point. Only with increased gas temperature can the stack temperature be reduced to 300 F. To relieve this problem, a two-pressure steam cycle is used. As shown in Figure [illegible], by adding a second low-pressure cycle it is possible to utilize the exhaust heat regardless of temperature. This is done at the expense of steam cycle performance with the efficiency being about 30 percent.

This basic difference in philosophy between operation of a conventional steam cycle and a COGAS steam cycle may require further clarification. In the conventional steam station, the use of regenerative feedwater heating results

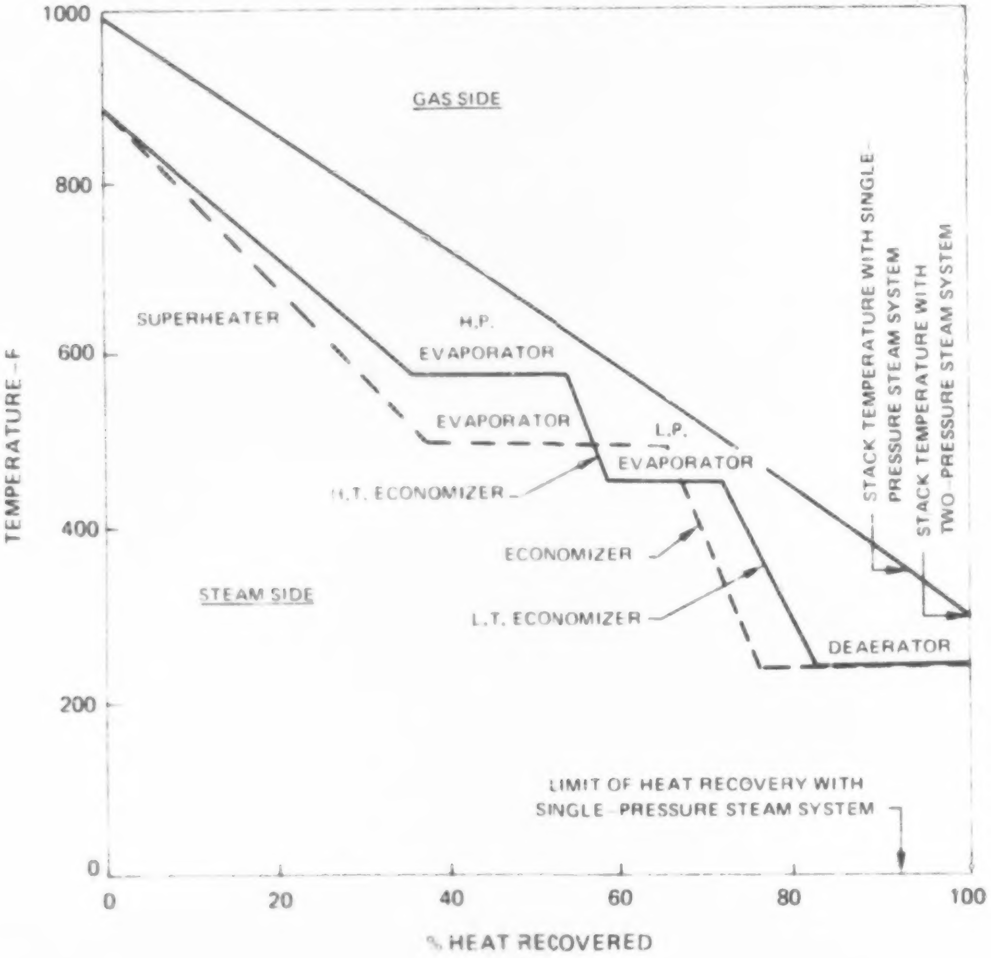
FIG. 22

TEMPERATURE DISTRIBUTION IN WASTE HEAT RECOVERY BOILER FOR OIL–FIRED COGAS STATION

MID–1970'S TECHNOLOGY

TURBINE INLET TEMPERATURE = 2200 F

COMPRESSOR PRESSURE RATIO = 16



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in exhaust gas temperature leaving the economizer section of the boiler in excess of 500 F and as high as 700 F. The remaining heat (above 300 F) is removed in an air preheater and, in effect, returned to the boiler. In the COGAS system it is not possible to use the stack gas in an air preheater since this would be harmful to gas turbine performance, increasing compressor power and raising compressor operating temperatures. The only means of reducing stack gas temperature to 300 F is to use the stack gas in an economizer. As a result, regenerative feedwater heating must be abondoned (at least above 250 F) and the steam cycle efficiency is decreased. However, since the feedwater heating is now provided by heat that would have been rejected up the stack, and also the steam that would have been used for that feed heating can now be expanded in a turbine, the net effect is an improvement in the overall boiler/steam cycle performance.

The large increase that is possible in steam cycle performance due to the foregoing is sometimes used to justify the use of supplementary firing. For a distillate fueled system, the effects of supplementary firing are shown in Figure 23. It can be seen that at lower turbine inlet temperatures a small improvement in efficiency is possible, but it is at the expense of added complexity in both the boiler and steam cycle. A similar study was conducted for a gasified coal system and it was shown that the gains due to supplementary firing were even less than for the distillate system. This is discussed in a later section, but it is the basis for considering only waste-heat-recovery-type COGAS systems.

COGAS Performance

To establish a performance base for the integrated gasified coal/COGAS stations, a series of calculations were made for a distillate fuel fired system. For each of the five combinations of materials and cooling techniques mentioned earlier, performance was evaluated over a temperature range of 2200 to 2900 F and over pressure ratios of interest. The results are presented in Figures 24 through 28. In all cases the steam cycle was optimized to give a stack temperature of 300 F while maintaining a steam temperature of 100 F less than turbine exhaust to a maximum of 1000 F. The steam cycle condenser pressure is 4 in. Hg. abs. With higher gas turbine exhaust temperatures (over 1200 F) a single pressure, non-reheat steam cycle is used since those conditions permit a 300 F stack temperature. Boiler supplementary firing was not used. In general, the peak efficiency point occurs with a steam temperature slightly below 1000 F and with a two-pressure steam cycle.

It is interesting to compare simple-cycle and COGAS performance. Referring to Figures 20 and 24 at peak COGAS efficiency, it can be seen that the steam cycle adds about 13 points to the corresponding simple-cycle performance.

COGAS STATION PERFORMANCE WITH FIRED BOILER

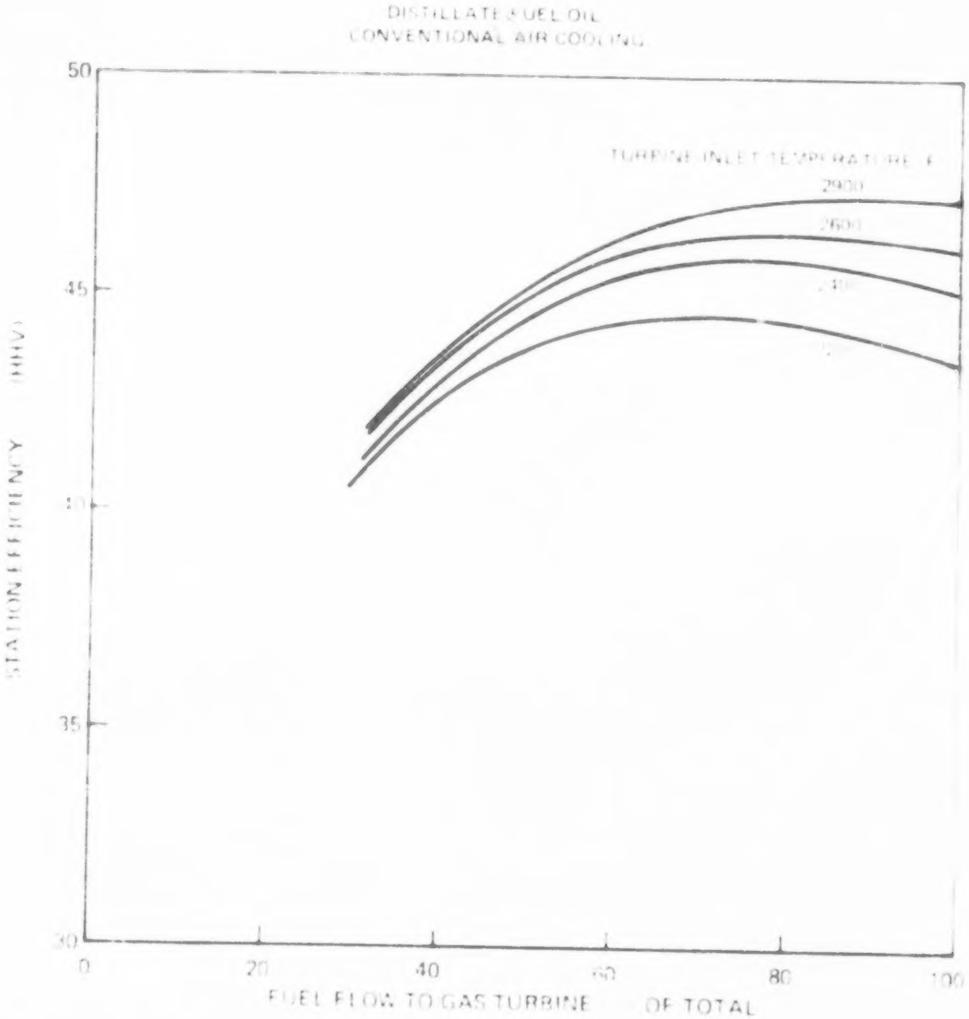

FIG. 24

COGAS STATION PERFORMANCE WITH CONVENTIONAL AIR COOLING

DISTILLATE FUEL OIL

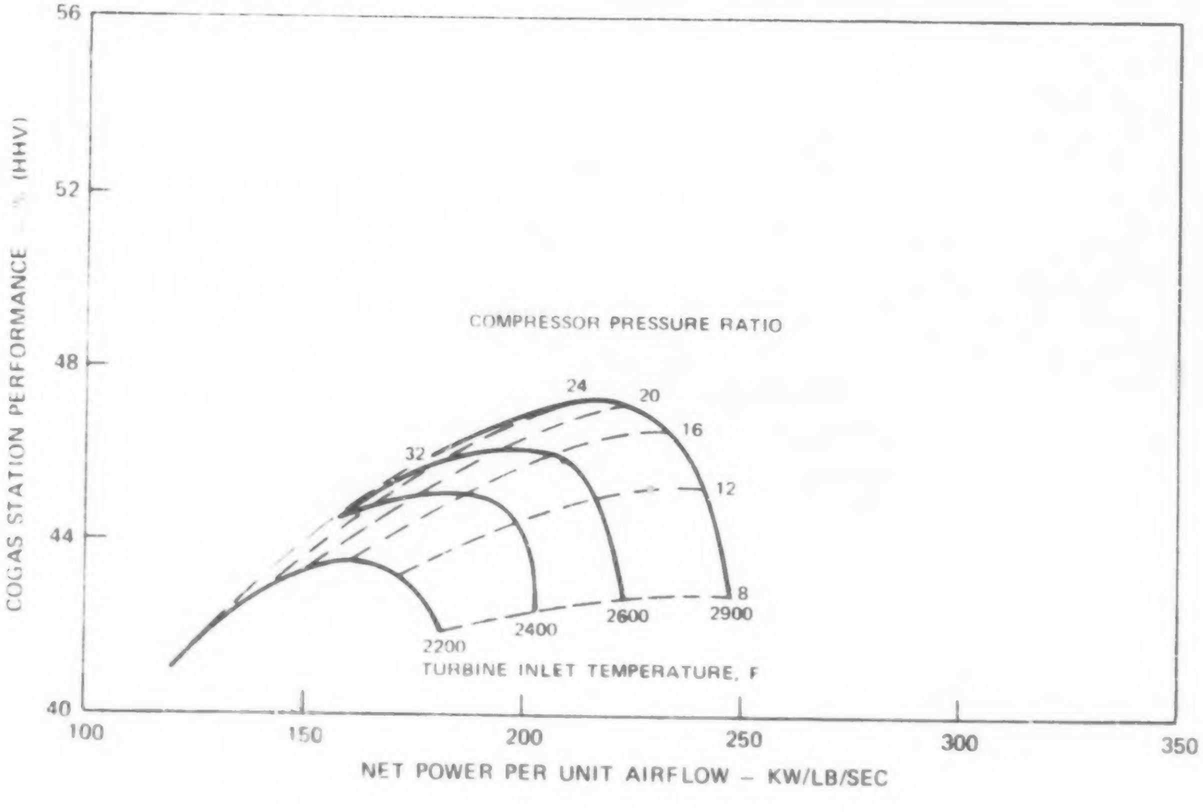

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FIG. 25

COGAS STATION PERFORMANCE WITH PRECOOLED AIR TO VANES

DISTILLATE FUEL OIL

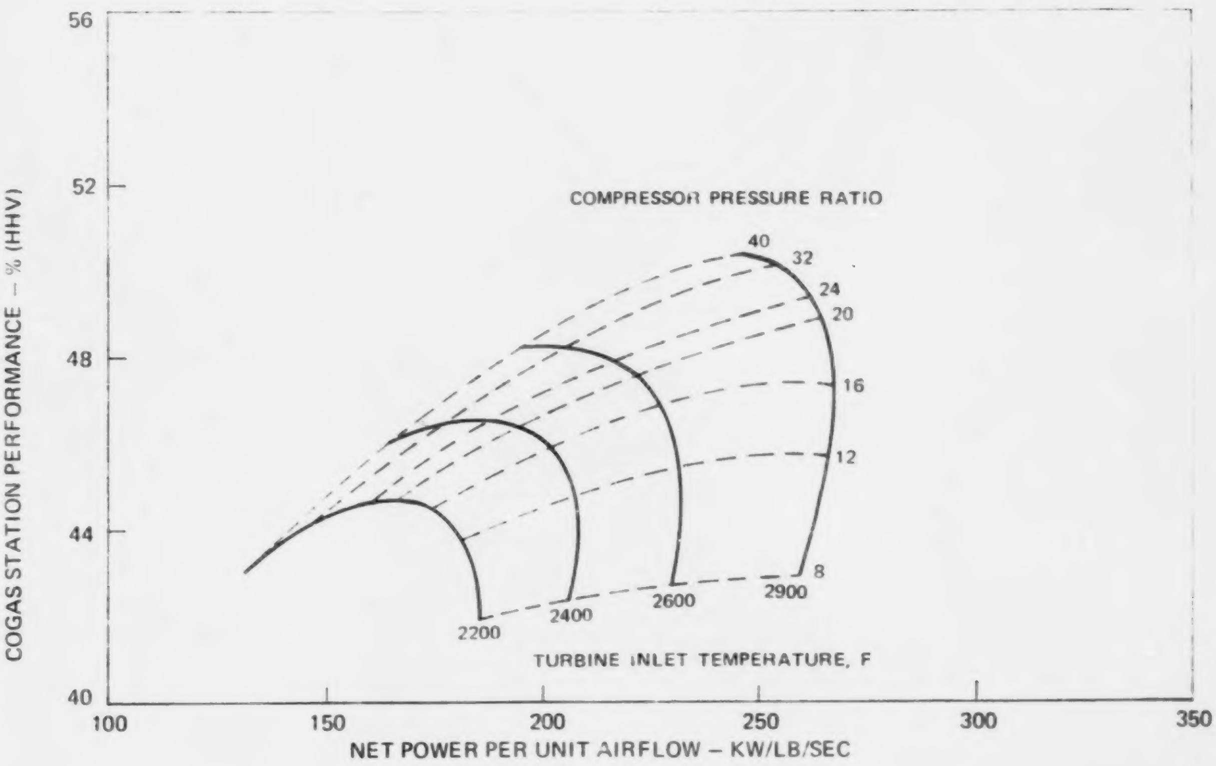

FIG. 26

COGAS STATION PERFORMANCE WITH CERAMIC VANES AND CONVENTIONAL COOLED BLADES

DISTILLATE FUEL OIL

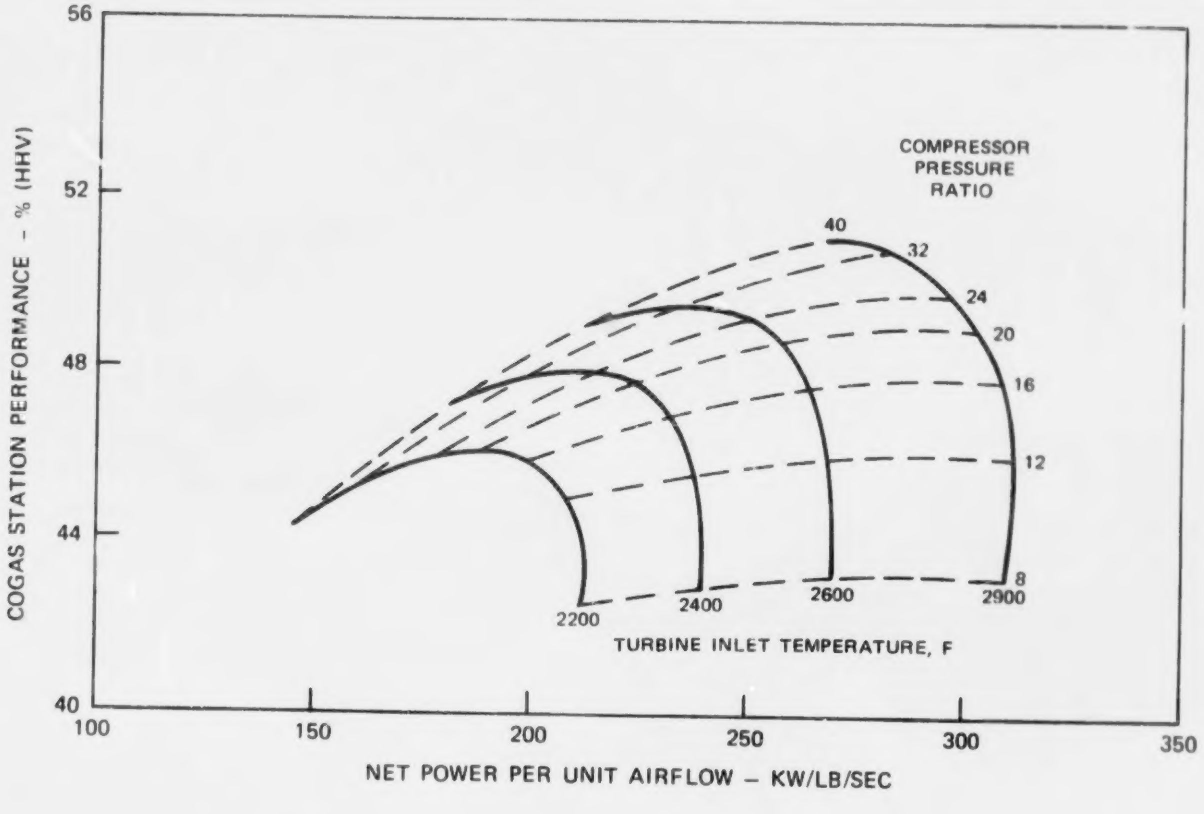

FIG. 27

COGAS STATION PERFORMANCE WITH CERAMIC VANES AND BLADES

DISTILLATE FUEL OIL

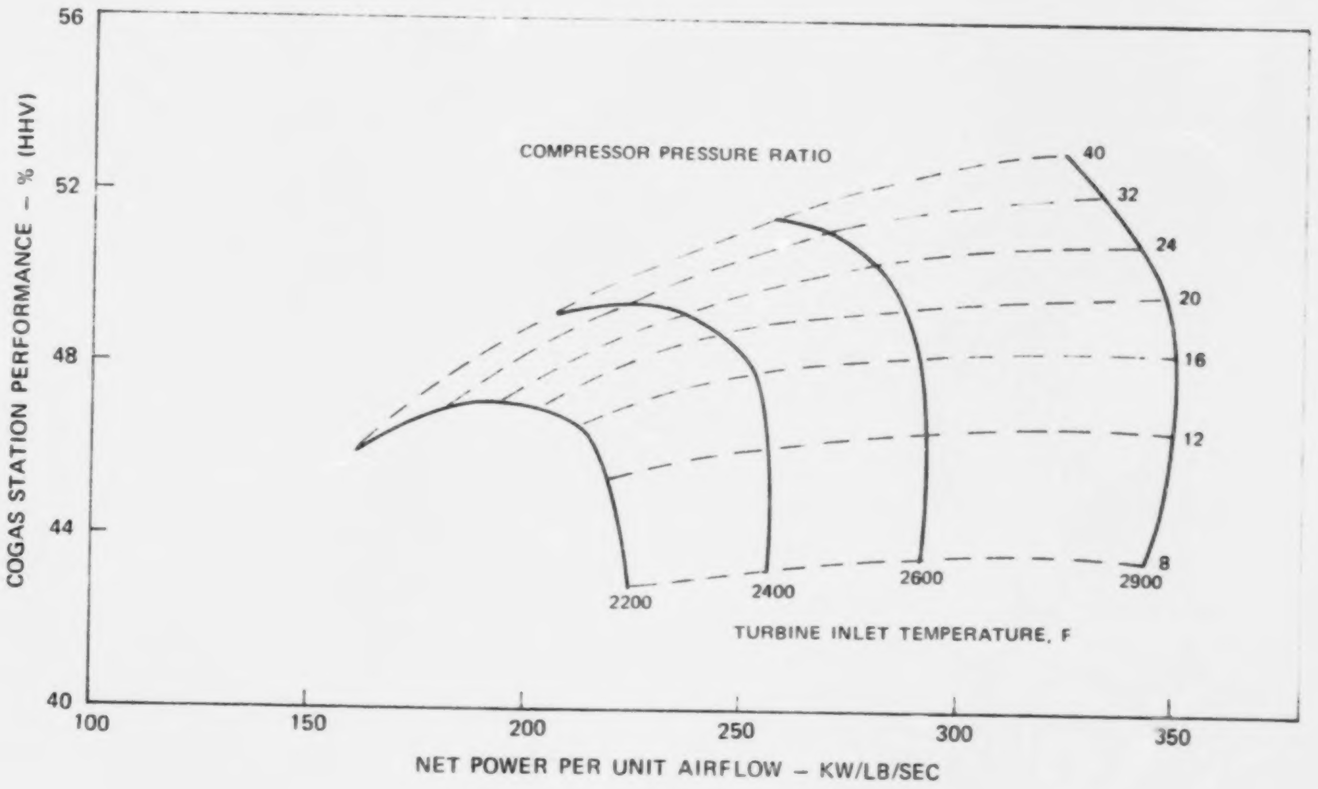

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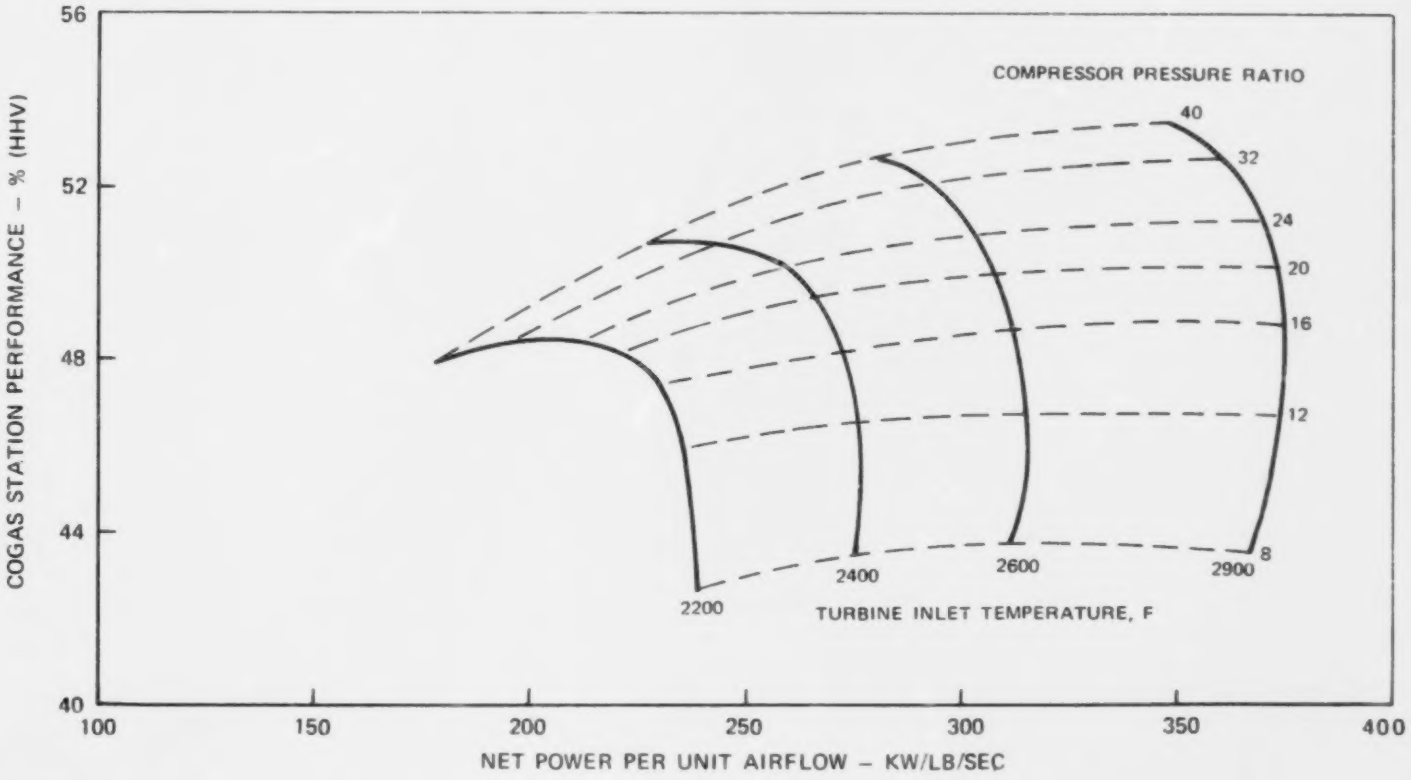
FIG. 28

COGAS STATION PERFORMANCE WITH ZERO BLEEDS AND LEAKAGE

DISTILLATE FUEL OIL



At the lower turbine inlet temperatures, simple-cycle efficiency continues to improve up to a pressure ratio of more than 24:1. In the combined system, such a high pressure ratio lowers exhaust temperature to a point where steam cycle performance falls off badly. Thus, the COGAS performance peaks at a much lower pressure ratio, reflecting a compromise between gas turbine and steam system performance. At higher turbine inlet temperatures, this effect tends to disappear. Peak simple-cycle efficiency then occurs at a pressure ratio corresponding to an exhaust temperature that permits good steam cycle performance. Thus, as turbine inlet temperature is increased, it not only permits a greater gas turbine output, but allows both gas turbine and steam cycles to perform at peak efficiency.

These same general characteristics apply to systems using a low-Btu gas. To permit evaluation of the integrated system, it was necessary to select a gas turbine pressure since it in turn sets gasifier and cleanup system operating pressure. Based upon the parametric studies, this was chosen to be the pressure ratio producing maximum specific power without a significant reduction in efficiency. For the conventionally cooled, 2200 F, first-generation system, a value of 16:1 was selected. A 24:1 pressure ratio was chosen for the second-generation, 2600 F engine with ceramic vanes and conventionally cooled blades.

Integrated System Performance Evaluation - The combination of gasifier, cleanup system, gas turbine and steam cycle give rise to a very large number of possible configurations, especially when the necessary auxiliaries such as heat exchangers, booster compressors. boilers, etc. are taken into account. In order to investigate many of these combinations, a versatile computer model was developed at UTRC under Corporate sponsorship which allows great flexibility in analyzing the integrated systems. The State-of-the-Art Performance Program (SOAPP) is basically a bookkeeping system that permits the user to assemble individual modules into an integrated power system (see Appendix A). A number of different configurations were used during the course of the study, the most important of which are shown in Figures 29 through 31.

The configuration shown in Figure 29 was used extensively in preliminary integration studies. The left column of components comprises the main elements of a waste-heat-recovery COGAS power system. The gas generator is a dual spool machine, i.e., separate high-pressure and low-pressure shafts operating at different speeds with provision for intercooling and for bleeding the necessary air for the gasifier. The main heat-recovery boiler also has provision for supplementary firing and the steam cycle can be selected to suit the exhaust gas temperature. In general, steam raised elsewhere in the system is used for process steam; however, provision is made to supplement the steam

FIG. 29

INTEGRATED COGAS/COAL GASIFICATION POWER STATION

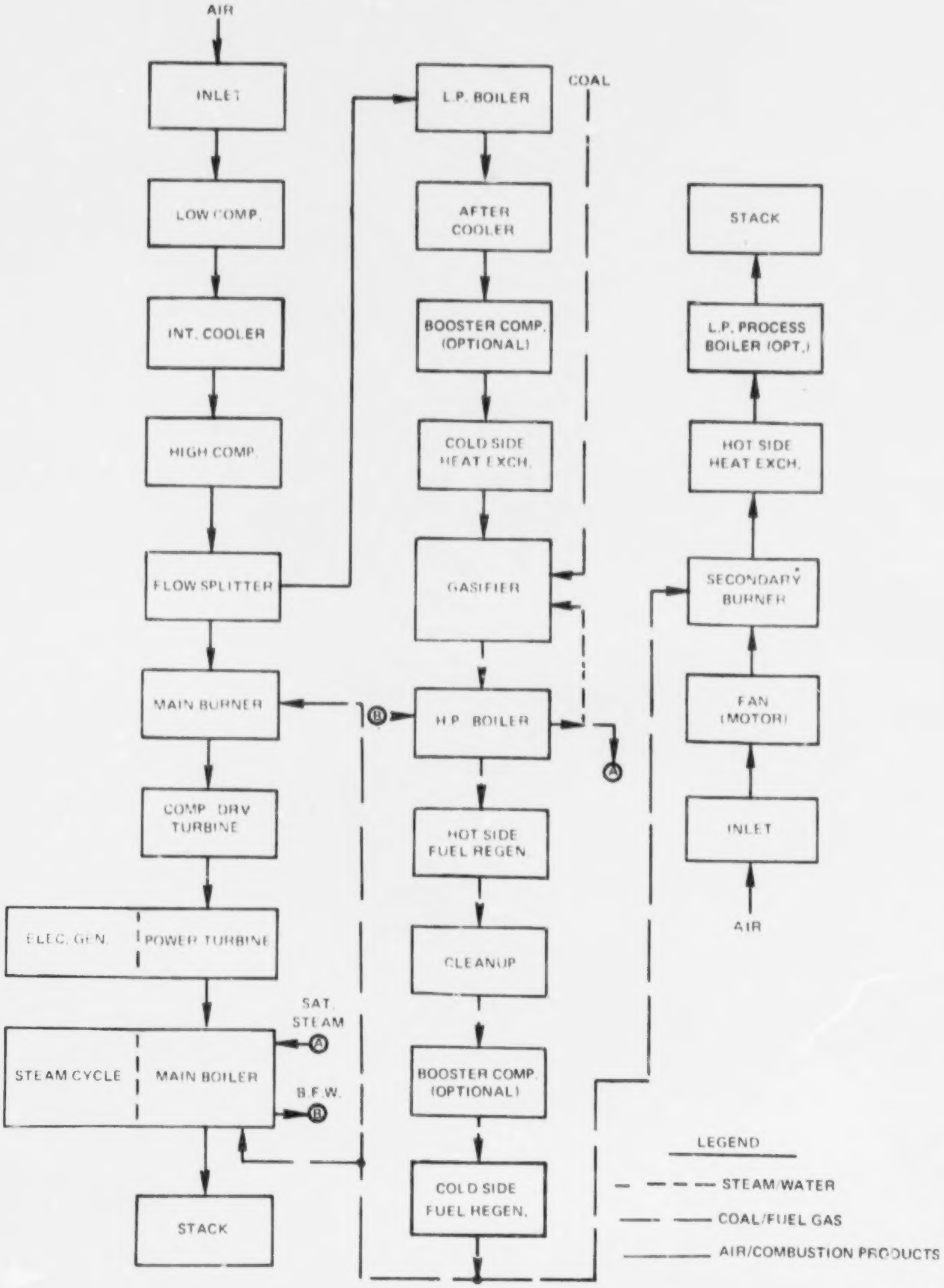

AIR
INLET
LOW COMP.
INT. COOLER
HIGH COMP.
FLOW SPLITTER
MAIN BURNER
COMP. DRV TURBINE
ELEC. GEN.
POWER TURBINE
SAT. STEAM
A
STEAM CYCLE
MAIN BOILER
B.F.W.
B
STACK
L.P. BOILER
COAL
AFTER COOLER
BOOSTER COMP. (OPTIONAL)
COLD SIDE HEAT EXCH.
GASIFIER
B
H.P. BOILER
A
HOT SIDE FUEL REGEN.
CLEANUP
BOOSTER COMP. (OPTIONAL)
COLD SIDE FUEL REGEN.
STACK
L.P. PROCESS BOILER (OPT.)
HOT SIDE HEAT EXCH.
SECONDARY BURNER
FAN (MOTOR)
INLET
AIR
LEGEND
STEAM/WATER
COAL/FUEL GAS
AIR/COMBUSTION PRODUCTS

FIG. 30

INTEGRATED COGAS/COAL GASIFICATION POWER STATION WITH LOW TEMPERATURE CLEANUP SYSTEMS

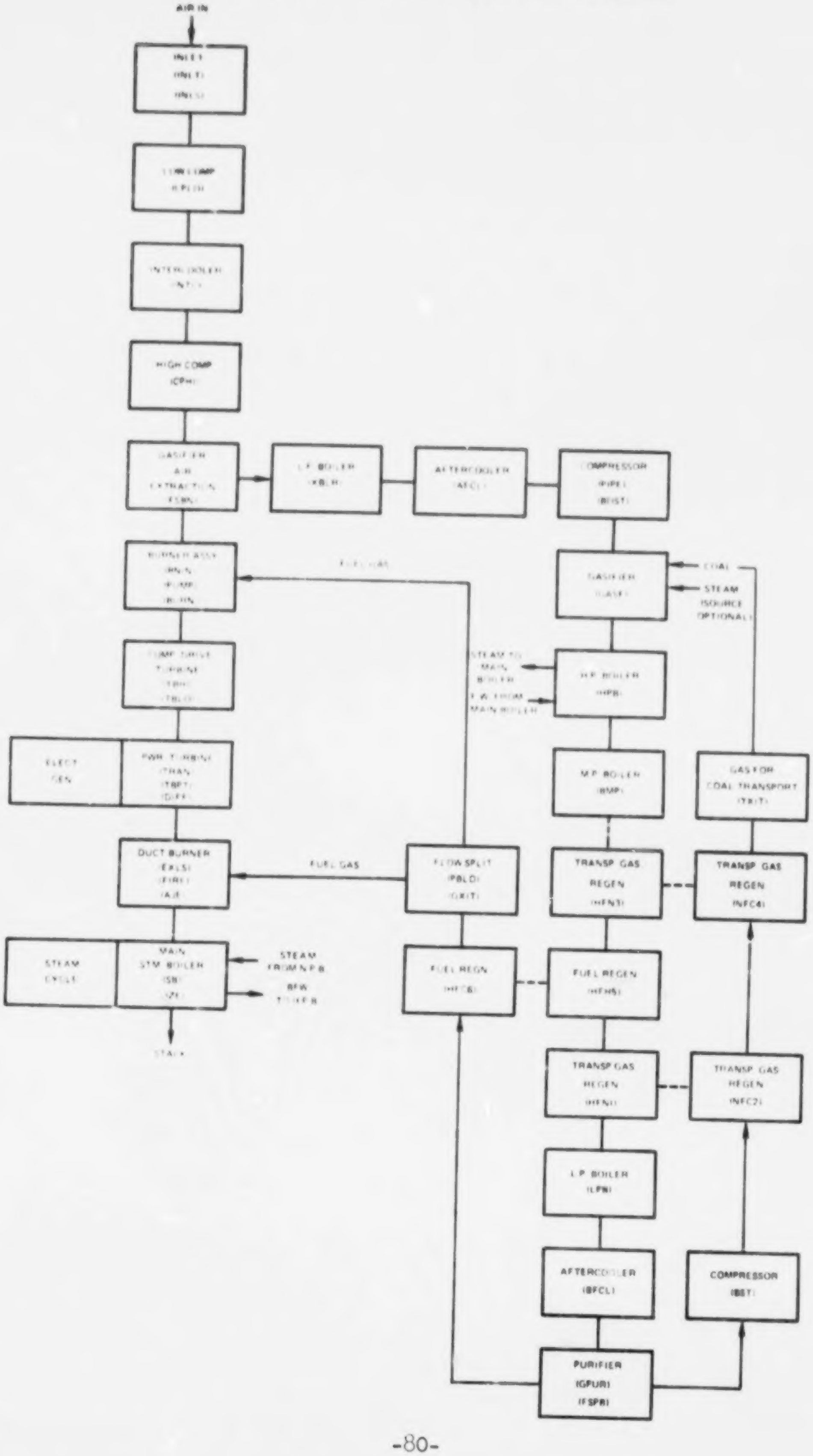
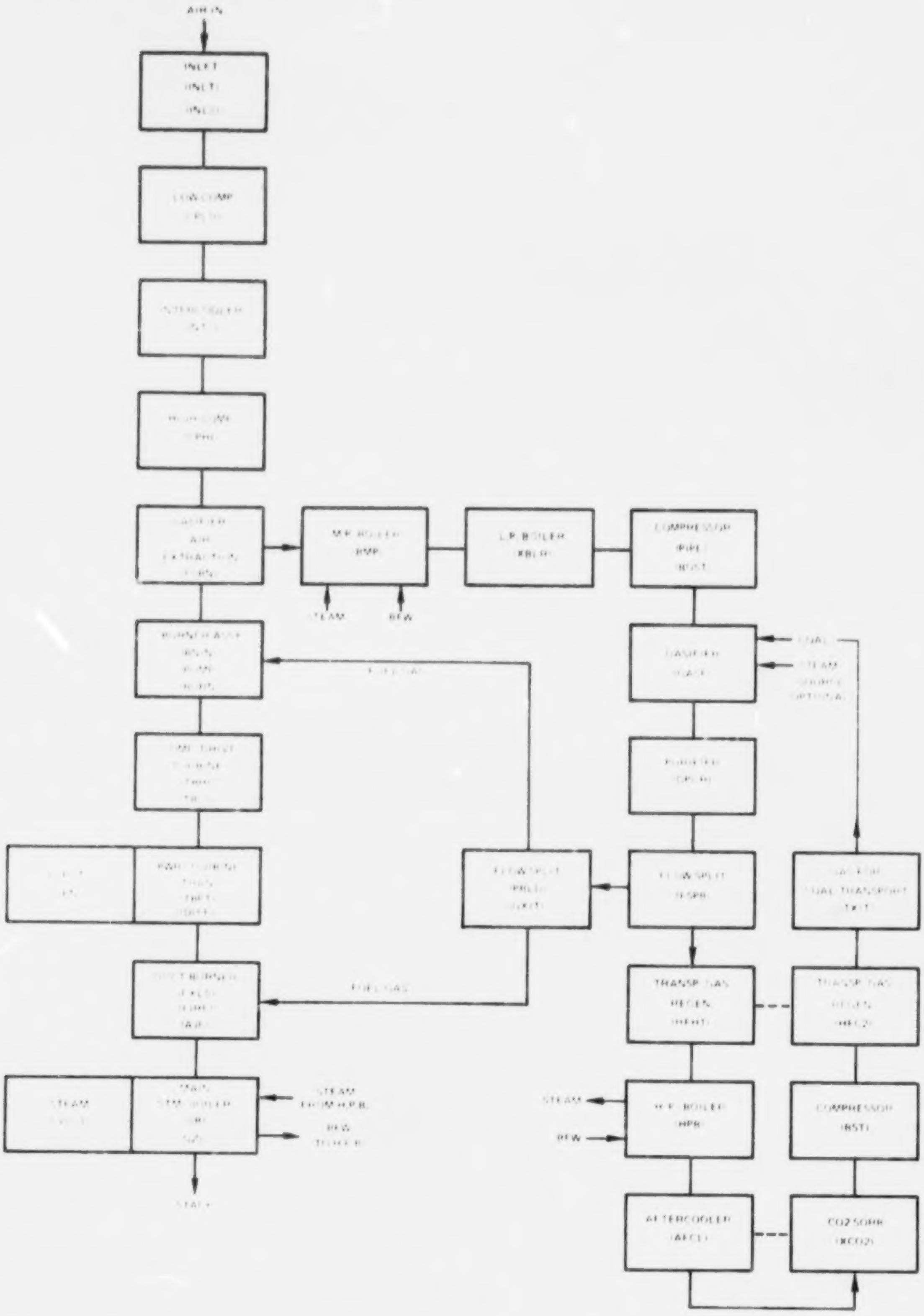

AIR IN
INLET
(INLT)
(INLS)
LOW COMP
(LPCO)
INTERCOOLER
(INTC)
HIGH COMP
(CPH)
GASIFIER
AIR
EXTRACTION
(ESBN)
L.P. BOILER
(XBLR)
AFTERCOOLER
(AFCL)
COMPRESSOR
(PIPE)
(BIST)
BURNER ASSY
(BNIN)
(PUMP)
(BURN)
FUEL GAS
GASIFIER
(GASF)
COAL
STEAM
(SOURCE
OPTIONAL)
PUMP DRIVE
TURBINE
(TBH)
(TBLD)
STEAM TO
MAIN
BOILER
F.W. FROM
MAIN BOILER
H.P. BOILER
(HPB)
ELECT
GEN
PWR TURBINE
(TRAN)
(TBPT)
(DIFF)
M.P. BOILER
(BMP)
GAS FOR
COAL TRANSPORT
(TRIT)
DUCT BURNER
(EXLS)
(FIRE)
(AJE)
FUEL GAS
FLOW SPLIT
(PBLD)
(GRIT)
TRANSP GAS
REGEN
(HFN3)
TRANSP GAS
REGEN
(NFC4)
STEAM
CYCLE
MAIN
STM BOILER
(SB)
(ZI)
STEAM
FROM N.P.B.
BFW
TO H.P.B.
FUEL REGN
(HFC6)
FUEL REGEN
(HFN5)
STACK
TRANSP GAS
REGEN
(HFN1)
TRANSP GAS
REGEN
(NFC2)
L.P. BOILER
(LPB)
AFTERCOOLER
(BFCL)
COMPRESSOR
(BST)
PURIFIER
(GPUR)
(FSPR)

FIG. 31

INTEGRATED BCR/CONSOL GASIFICATION POWER STATION

HIGH TEMPERATURE CLEANUP SYSTEMS


AIR IN
INLET
LOW COMP
INTERCOOLER
HIGH COMP
GASIFIER AIR EXTRACTION
M.P. BOILER
(BMP)
STEAM
BFW
L.P. BOILER
(XBLP)
COMPRESSOR
(PIPE)
(BST)
BURNER ASSY
FUEL GAS
GASIFIER
(GASF)
COAL
STEAM SOURCE OPTIONAL
FLOW SPLIT
FLOW SPLIT
(FSPR)
GAS FOR COAL TRANSPORT
(TXT)
PWR TURBINE
POST BURNER
FUEL GAS
TRANSP. GAS REGEN.
(REHT)
TRANSP. GAS REGEN.
(HEC2)
STEAM
MAIN STM BOILER
STEAM FROM H.P.B.
BFW
STEAM
BFW
H.P. BOILER
(HPB)
COMPRESSOR
(BST)
STACK
AFTERCOOLER
(AFCL)
CO2 SORB
(XCO2)

generated in the main boiler with steam raised in the gasifier-cleanup part of the system. The center column represents the gasifier and cleanup system. Air at compressor outlet conditions is cooled in a boiler and aftercooler to a temperature that is dependent on boost compressor requirements. If necessary, an air preheater can be supplied with heat from the secondary burner in the right hand column. The steam consumed by the gasifier is raised in either the main or H.P. boiler. Gasifier jacket heat is used for feedwater and steam generation. Heat from the gasifier effluent can be recovered in the H.P. boiler and in a fuel gas regenerator when used with a low-temperature cleanup process. The requirements for steam at various pressures in the cleanup system are used to control steam generation rates and to determine net clean fuel gas and power output.

In general, any of the modules can be rendered inoperative resulting in a great degree of flexibility within a single configuration. For example, Figure 30 shows the configuration used to represent the BCR/low-temperature cleanup and both the low- and high-temperature cleanup in conjunction with the BuMines gasifier. In the case of the BuMines systems, many of the modules are not used because of the tar in the gasifier outlet stream which prevents heat recovery from that stream. In this configuration, additional boilers, heat exchangers and a boost compressor for the BCR transport gas stream (right-hand column) were added. It was more attractive to regenerate the transport gas as a means of satisfying the gasifier heat balance than either raising steam with a higher degree of superheat than that from the main stream cycle or increasing the gasifier air/fuel ratio.

The BCR/high-temperature cleanup system is represented by Figure 31. The differences between this and the low-temperature case are due to the higher temperature cleanup (no heat recovery) and the need to remove CO_2 from the transport gas stream for use in the regeneration of some of the high-temperature absorbers. The system is designed to raise process steam in the L.P. boiler with steam from both the M.P. and H.P. boiler used to supplement the main steam cycle.

Integrated System Studies - Prior to selection of the representative configurations for detailed evaluation, a number of general studies were performed to determine trends that will be generally applicable to all systems. Much of this work deals with the utilization of available process heat and points up the value of regeneration in the fuel stream and the advantages that accrue from using energy at the full cycle efficiency. Wherever possible, it is desirable to extract process heat only after expanding the working fluid through the gas turbine. Because of the large amount of process heat used in the gasification and cleanup systems, these factors are magnified as compared to a distillate-fired COGAS system.

The analyses discussed in the following paragraphs were based on preliminary data for gasification and were intended to identify the trends in performance due to variations in operating characteristics. Thus the absolute values of performance may differ from the final values obtained for the detailed analyses of selected systems.

Boiler Firing Studies - Using preliminary data for a BCR type gasifier with a representative low-temperature cleanup system (Selexol), a series of performance analyses were conducted. One part of that series deals with the use of clean product gas to increase steam boiler gas inlet temperature. As discussed previously the evaluation of fired vs. unfired boilers is relatively complex and cannot be answered simply. The flow diagram used in these calculations is basically that shown in Figure 29 with the secondary burner stream omitted. It was unnecessary to aftercool following the M.P. boiler and the boost compressor is used to raise the air temperature to satisfy gasifier inlet air temperature requirements of 800 F. The performance of a first-generation COGAS system with various steam cycles is presented in Figure 32. With no boiler firing and a two-pressure steam cycle, peak efficiency occurs at an exhaust gas temperature to the boiler of about 925 F. This corresponds to a pressure ratio of 16:1. Steam [illegible] for the other cycles considered are noted on the figure. In all cases, supplementary firing was required to bring the boiler gas temperature to a high enough level to maintain the designated steam conditions and to fully utilize the heat in the gas stream by maintaining a 300 F stack gas temperature. At boiler gas temperatures lower than this required level the stack temperature would exceed 300 F at which point overall efficiency falls off quite rapidly. It is apparent that because of the supplementary fuel that must be burned to utilize the high-performance steam cycles, the resulting station efficiency is no better than when a relatively simple, low-cost steam cycle is used. As a result of this analysis, integration efforts were concentrated on the use of simple steam cycles and unfired boilers.

Steam Conditions - Using the above system configuration, the effect of varying steam cycle pressure was investigated. The net efficiency for the station having an unfired boiler with a two-pressure steam system with the high pressure set at 1250 psi is shown in Figure 32 to be 33.7 percent. Performance at this condition was selected as a base-line point for comparing other cycle variations. The cycle parameters corresponding to that point are given in Table 10. The effect of changing gas turbine compressor pressure ratio and the throttle pressure levels of the high-pressure portion of the steam cycle from the reference level is also shown in Figure 33. Station efficiency is relatively insensitive to compressor pressure ratio about the reference point. Similarly, efficiency is relatively insensitive to steam pressure as it is increased from 1250 to 1800 psi. The 1800 psia level represents an upper limit

FIG. 32

EFFECT OF STEAM CYCLE ON STATION PERFORMANCE

TURBINE INLET TEMPERATURE = 2200F
COMPRESSOR PRESSURE RATIO = 16, UNLESS NOTED
SINGLE PRESSURE STEAM SYSTEM, UNLESS NOTED

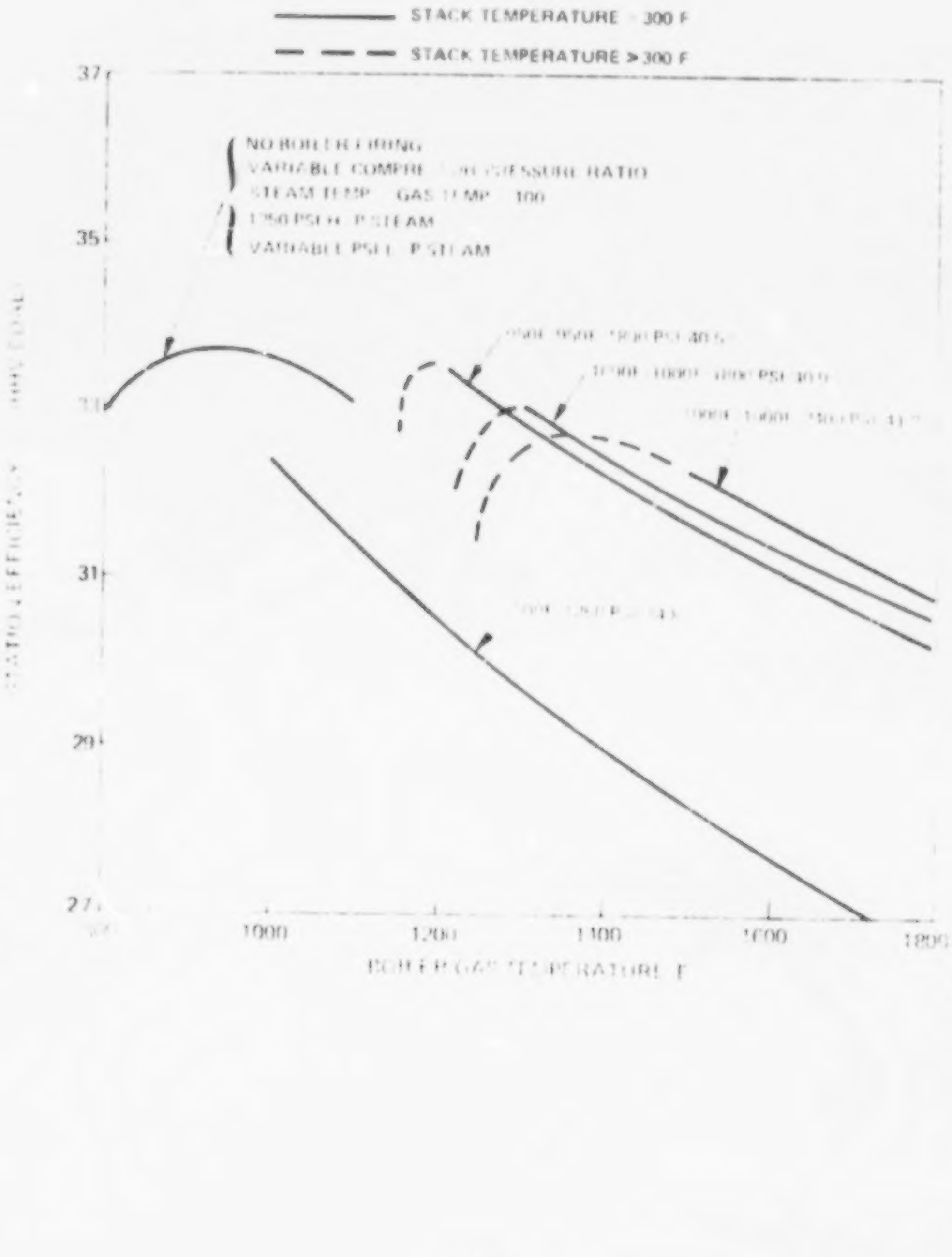


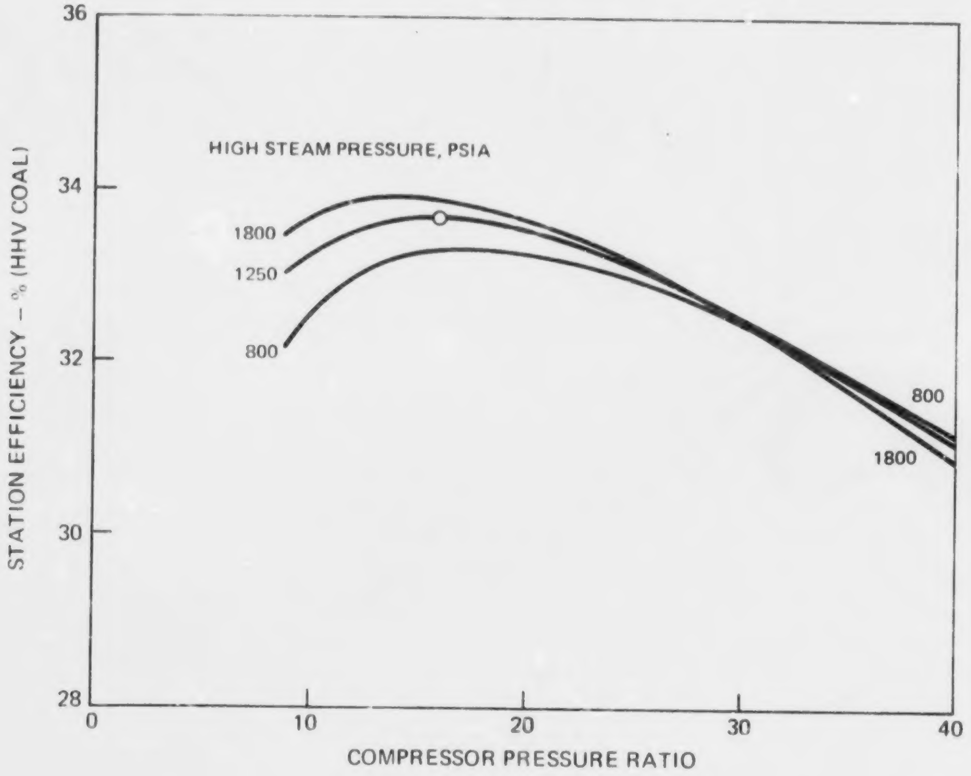
Table 10

DESIGN CHARACTERISTICS FOR REFERENCE INTEGRATED COGAS POWER STATION

Gas Turbine	
turbine inlet temperature	2200 F
compressor pressure ratio	16
air bleed for gasifier	11.8%
Steam Cycle (Unfired Boiler)	
steam temperature (non-reheat)	827 F
superheater pinch temperature difference	100 F
high steam pressure	1250 psia
h-p evaporator pinch temperature difference	50 F
low steam pressure	334 psia
l-p evaporator pinch temperature difference	50 F
economizer pinch temperature difference	50 F
stack temperature	300 F
Gasifier and Cleanup Systems	
BCR gasifier exit temperature	1800 F
air supply temperature (no preheat)	800 F
steam pressure to gasifier	1250 psia (sat)
temperature of feedwater to gasifier h-p boiler	250 F
booster compressor pressure ratio	2
Selexol operating temperature	95 F
fuel regenerator effectiveness	0.9
regenerator exit temperature (hot side)	300 F
fuel supply temperature (regenerator exit, cold side)	677 F
fuel sensible heat	12 Btu/SCF
fuel higher heating value	157 Btu/SCF
Integrated Station	
auxiliary consumption	6%

INTEGRATED COGAS/BCR GASIFICATION STATION PERFORMANCE

TURBINE INLET TEMPERATURE = 2200 F
TWO–PRESSURE STEAM SYSTEM
STACK TEMPERATURE = 300 F
FUEL REGENERATOR EFFECTIVENESS = 0.9
UTILIZE ALL SENSIBLE HEAT ABOVE 300 F



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and higher pressures cannot be used without reheat or excessive moisture would form in the low-pressure section of the turbine. Because of the relatively small influence of steam pressure, the 1250 psi level was selected as a standard for the integrated systems. Thus, if it appears desirable to raise steam pressure, it would be possible to re-evaluate any of the selected configurations without changing gas turbine compressor pressure ratio or gasifier and cleanup system operating pressure levels.

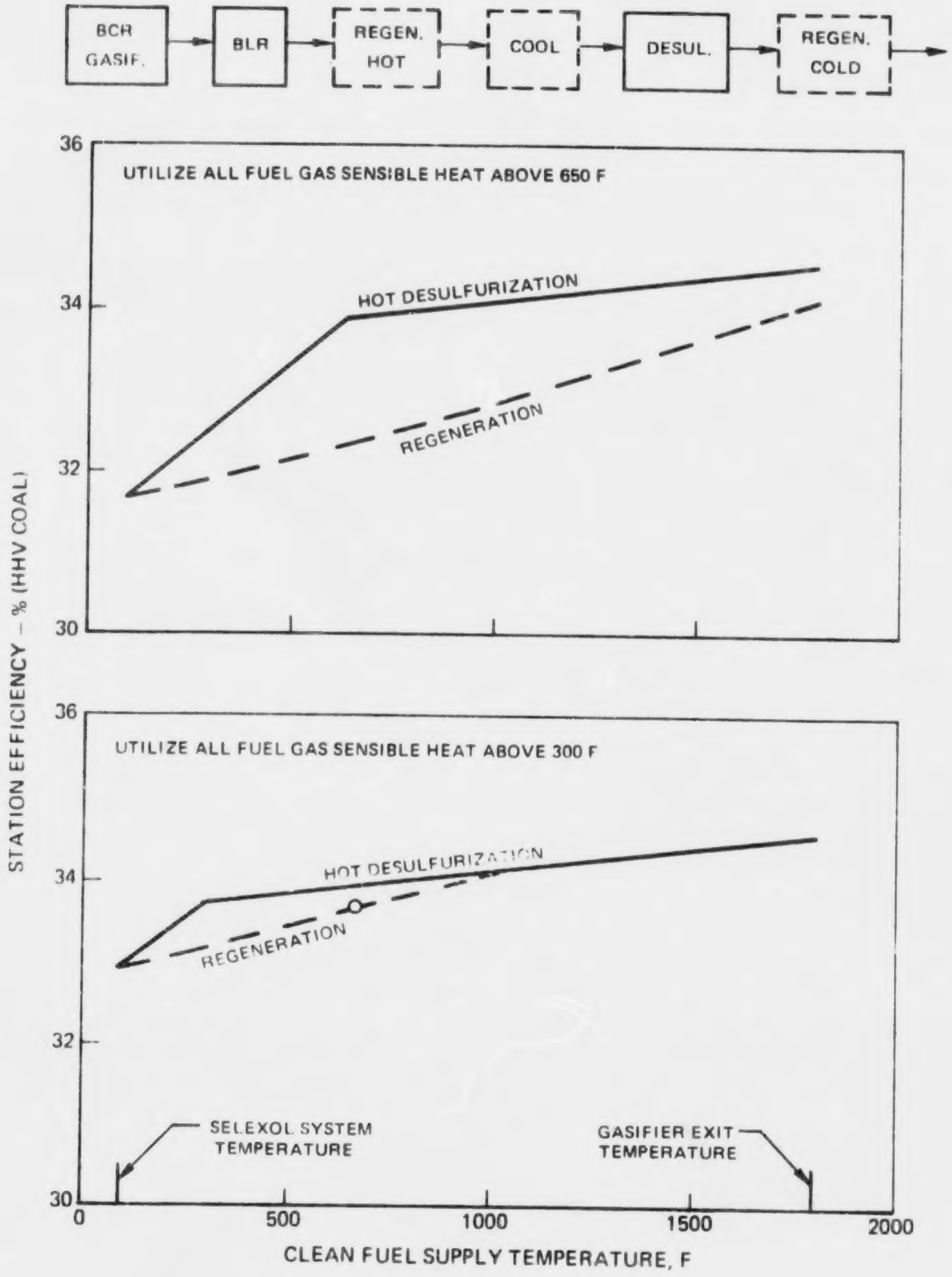
Fuel-Gas Temperature Effects - For the base-line system the effects of high-temperature cleanup and fuel-gas regeneration were evaluated. The performance improvements possible by using elevated fuel supply temperatures are shown in Figure 34. Two methods were investigated for achieving elevated fuel temperature: fuel regeneration (using hot, dirty fuel to reheat clean fuel from the low-temperature system) and hot desulfurization (using a fictitious high-temperature cleanup system which removes only H_2S, CO_2, NH_3, and particulate matter). The performance of the reference station with fuel regeneration to 677 F is noted in the lower part of Figure 34. The performance improvement depends heavily on the temperature level below which sensible heat in the hot fuel gas stream is not recovered (i.e., minimum temperature at exit of boiler or hot side of fuel regenerator). Recovery of this low-temperature heat is made more difficult by the presence of water vapor, sulfur compounds and ammonia in the dirty gas which results in a mass flow rate that is higher than for the clean gas on the cold side of the regenerator. As a result, the temperature drop on the hot side of the regenerator is significantly less than on the cold side. One of the undesirable aspects of a low-temperature cleanup system is the need to condense most of the water vapor in the dirty gas stream. This will result in the need for costly heat exchange equipment to withstand the weak acids which will be present. However, since no alternative is apparent it appears desirable to utilize as much of the heat in the dirty gas stream as is possible to ease the duty of the regenerator.

The foregoing analysis placed no restrictions on regenerator performance or design characteristics. A highly effective regenerator will result in excessive size and cost. Also, one that is capable of operating at temperatures up to 1000 F will require more expensive materials. To evaluate the desirability of regeneration to 1000 F, a cost study was performed for the BCR-Selexol configuration. Figure 35 shows the effect of regeneration on overall efficiency and on the heat exchanger area requirements as a function of the regenerator outlet temperature. Because of the rapidly increasing area, the high-temperature exchanger shows a capital cost increment due to both materials and size. Based on the area requirements of Figure 35, a comparison of regenerator cost was made between a 750 F outlet temperature and a 1000 F outlet temperature. This is presented in Table 11. Equipment costs

FIG. 34

PERFORMANCE IMPROVEMENT DUE TO ELEVATED FUEL SUPPLY TEMPERATURE

TURBINE INLET TEMPERATURE = 2200 F
COMPRESSOR PRESSURE RATIO = 16
TWO-PRESSURE STEAM SYSTEM
HIGH STEAM PRESSURE = 1250 PSIA
STACK TEMPERATURE = 300 F



EFFECT OF FUEL TEMPERATURE

BCR/SELOXOL SYSTEM

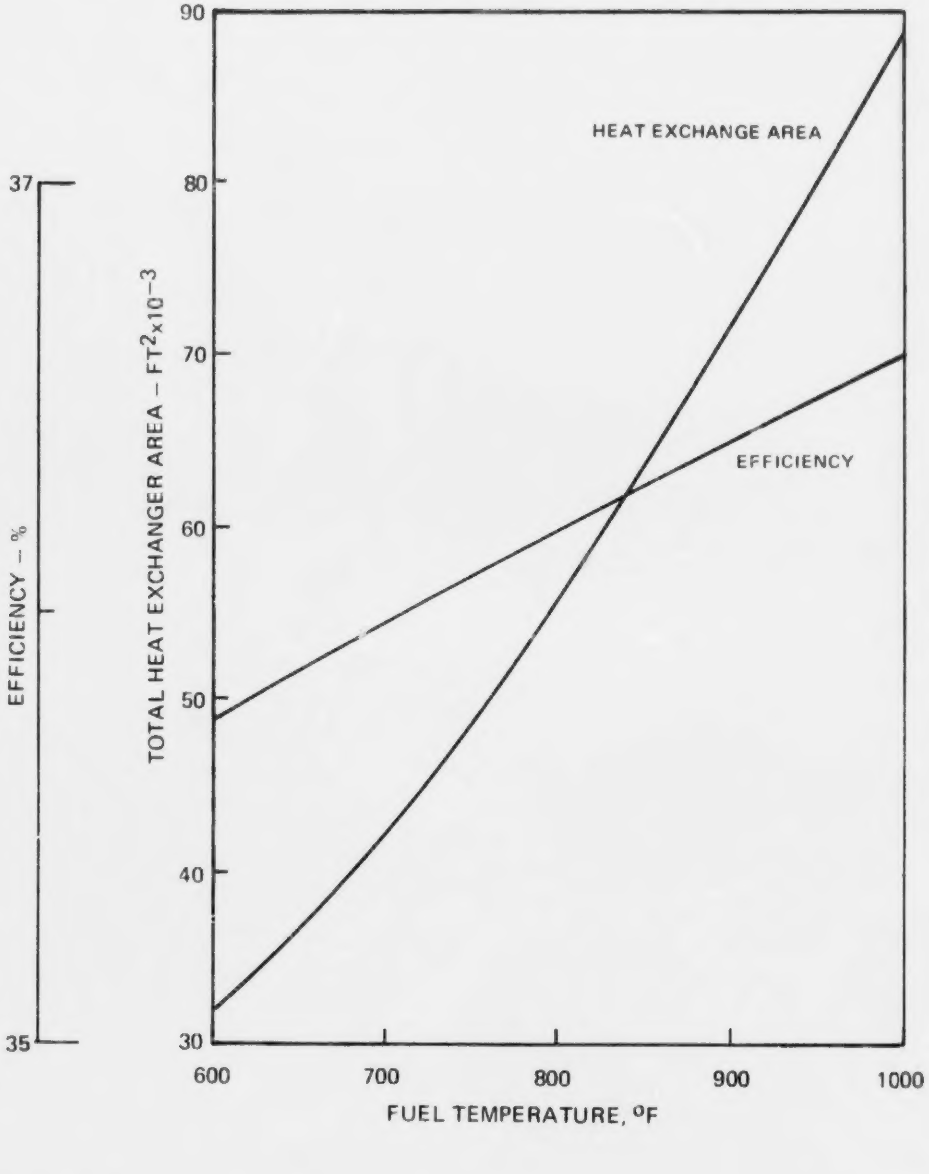


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Table 11

LCR-SELEXOL FUEL GAS REGENERATOR COSTS

Fuel Supply Temp - F	1000	750
Required HX Area - ft^2	88,500	48,500
Exchanger Cost-$	2.43×10^6	0.65×10^6
Total D & E cost-$	4.38×10^6	1.53×10^6

were estimated to be \$17.50/ft^2 for the low-temperature unit and \$27.50/ft^2 for the 1000 F unit. These are increased by a 1.8 factor to arrive at total erected cost. In the high-temperature case, 18-8 stainless steel and chrome alloy steel are employed for the tube and shell materials. While low alloy carbon steel can be used for the shell in the low-temperature case, stainless steel would still be required for the tube service due to the H_2-H_2S environment at the operating temperature. The resultant cost differential is still quite substantial due to the significantly larger area required for the high-temperature regeneration. At the net plant output, the incremental capital cost per point improvement in efficiency is \$5.72/kw. This can be related to a fuel cost using Figure 36 where yearly fuel savings due to improved efficiency have been related to an incremental capital investment. The high temperature regeneration is justified for fuel costs in excess of 60¢/MMBtu. Therefore, the BCR/Selexol system includes fuel regeneration to 1000 F.

Gasifier Heat Balance - The initial performance estimates for the BCR-type gasifier assumed the use of steam at 970 F which is about 150 F higher than would be supplied to the steam turbine. With the configuration of Figure 29 the alternative means of satisfying the gasifier heat balance were evaluated and generally found to be undesirable because of the relatively large penalties that resulted from the need to provide this small amount of additional heat. As an example, a decrease in efficiency of between 0.6 and 0.9 points results if the air preheater is used to provide the additional heat by raising inlet air temperature. A summary of the methods considered is presented in Table 12. It is interesting to note that the most costly (in terms of efficiency lost) means of providing the additional heat was to change the gasifier air-fuel ratio. The most desirable approach uses regeneration of the transport gas to a temperature of 694 F. This again serves to illustrate the value of regeneration in the fuel gas stream. As a result, the configurations shown in Figures 30 and 31 include provision for regeneration of the transport gas.

Effect of Pressure Ratio on Power Cost - The effect of gas turbine pressure ratio on the cost of power produced was evaluated for an integrated system based upon an entrained-flow residual oil gasifier. The characteristics of the system are presented in Table 13. The analysis of this system results in trends similar to coal-based systems. The constraints imposed on the power system were the same as those presented earlier with a two-pressure steam system optimized to produce a 300 F stack temperature. Steam temperature was set at 100 F below turbine exhaust with a maximum temperature of 1000 F. A low-temperature cleanup system was assumed. As pressure ratio was varied from 8 to 40 with a 2200 F turbine inlet temperature, efficiency varied from about 26.2 percent to 29.2 percent. An estimate of relative power costs is presented in Figure 37.

FIG. 36

CAPITAL INVESTMENT EQUIVALENT OF PLANT EFFICIENCY

17% YEARLY OWNING COST

0.7 LOAD FACTOR

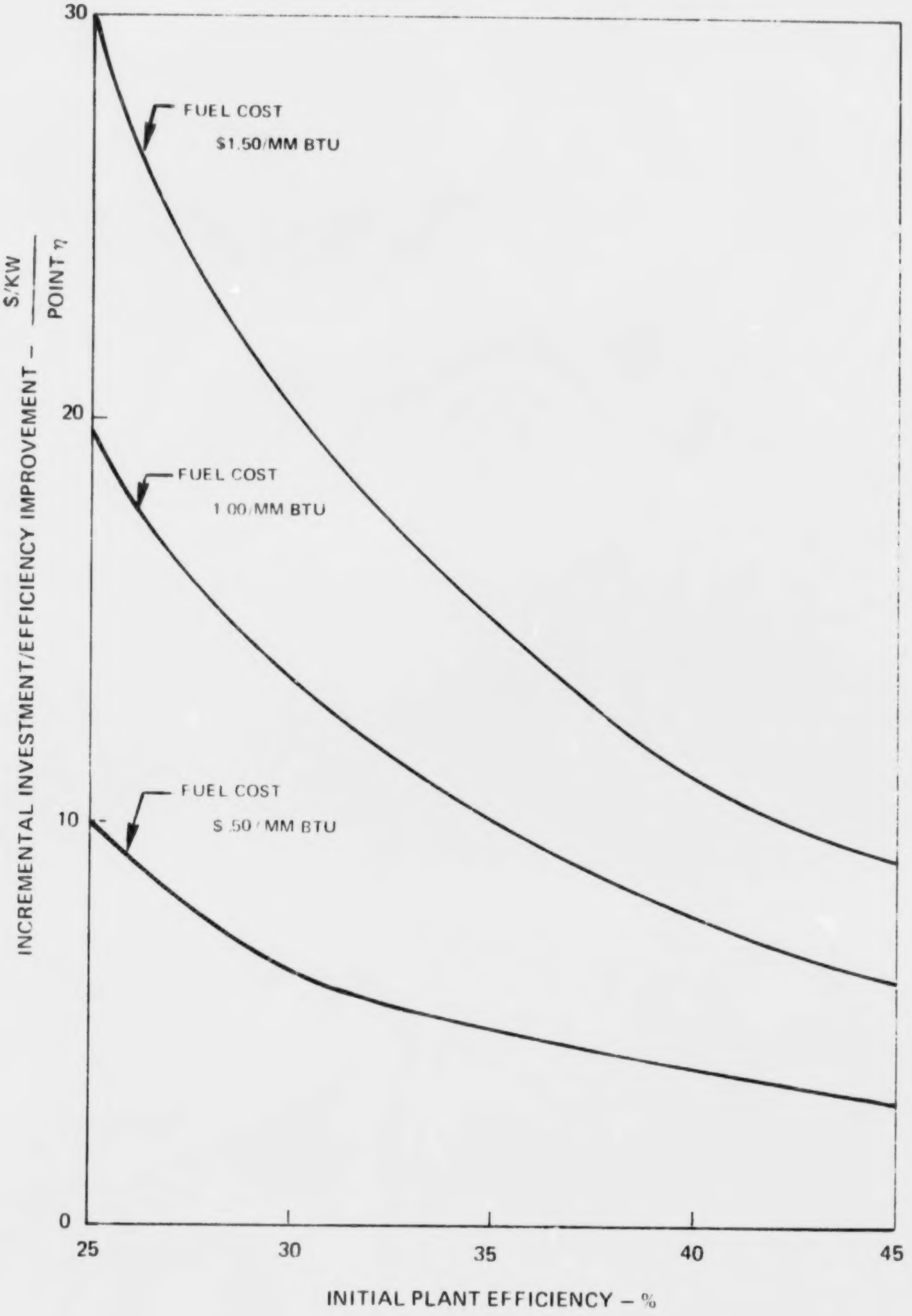

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Table 12

EFFECT OF STEAM EXTRACTION FOR GASIFIER ON STATION PERFORMANCE

Location of Steam Extraction	Steam Conditions T, F	Steam Conditions P, psia	Gasifier Air Temp, F	Exit Temp, F	Pressure Ratio	Transport Gas Temp, F	Efficiency Differential %	Comments
1. Gasifier HPB	584	1250	800	926	16.0	150	Base	Gasifier heat balance violated
2. Gasifier HPB	969	1250	800	926	16.0	150	-0.38	Only steam for gasifier superheated in HPB, high-risk-boiler
3. Main Boiler Superheater	526	1250	869	926	16.0	150	-0.59	Fired air preheater
4. Gasifier HPB	584 (Sat)	1250	1015	926	16.0	150	-0.87	Fired air preheater
5. Main Boiler Superheater	969	[illegible]	800	1069	[illegible]	150	-0.93	Special gas turbine design
6. Main Boiler Low-Pressure Evaporator	467 (Sat)	[illegible]	[illegible]	926	16.0	150	-1.10	Cannot maintain 300 F stack, fired air preheater
7. Main Boiler Superheater	950	900	800	1050	10.3	150	-1.26	Steam conditions per original BCR data, special gas turbine design
8. Main Boiler Superheater	969	1250	800	1069	16.0	150	-1.46	Fired boiler
9. Main Boiler Superheater	[illegible]	900	800	1050	16.0	150	-1.77	Steam conditions per original BCR data, fired boiler
10. Gasifier HPB	584 (Sat)	1250	800	926	16.0	150	-1.79	Base Case 1 with gasifier air/coal ratio altered to satisfy heat balance.
11. Gasifier HPB	969	1270	800	926	16.0	150	-2.55	Steam for gasifier superheated in special HPB
12. Main BLP	[illegible]	1250	800	926	16.0	694	-0.25	Additional gasifier heat obtained by preheating transport gas. Revised configuration also allows higher fuel temperature to engine.

Table 13

CHARACTERISTICS OF OIL GASIFICATION SYSTEM

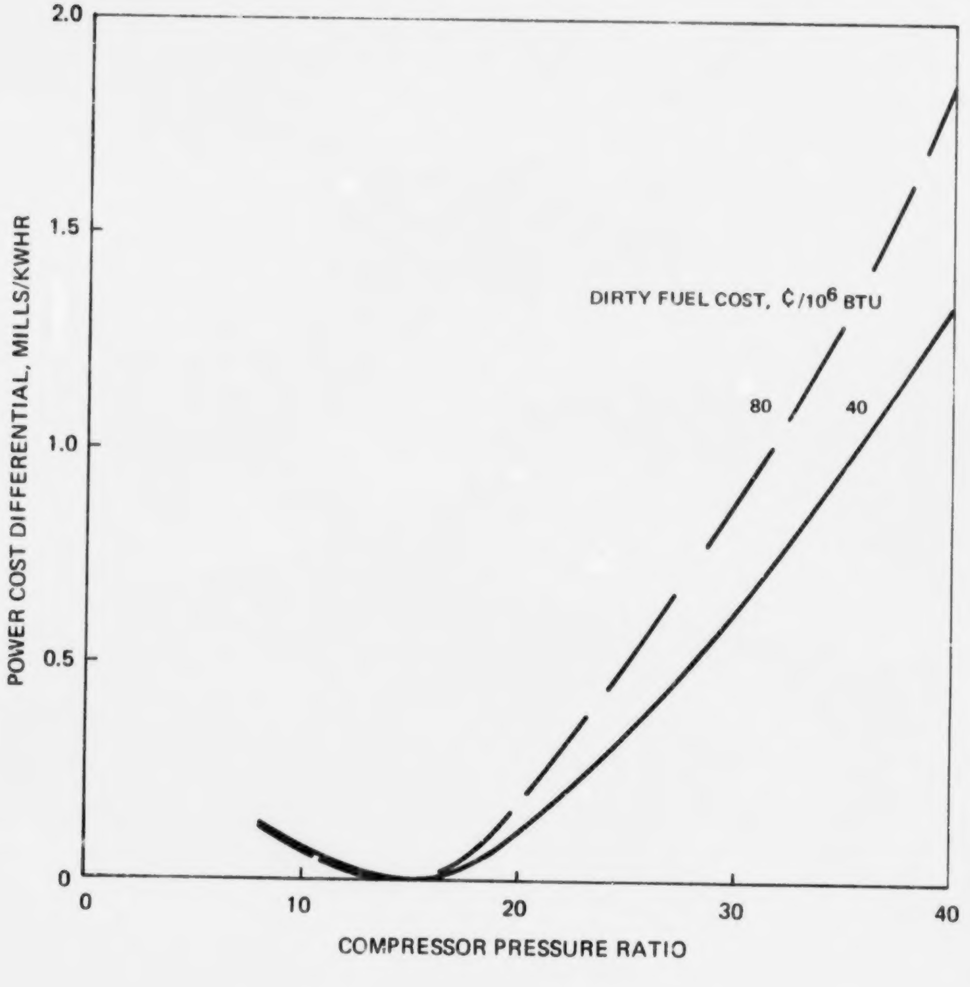
Gasifier:	Residual oil heating value	18,300 Btu/lb
	Air preheat temperature	955 F
	Air-to-oil ratio	6.39 lb air/lb oil
	Steam-to-oil ratio	0.15 lb steam/lb oil
	Exit gas temperature	2450 F
	Pressure drop	50 psia
Boiler	Steam pressure (saturated)	1250 psia
	Exit gas temperature	650 F
	Pressure drop	40 psia
Cleanup (Selexol):	Operating temperature	100 F
	Fuel gas consumption	1%
	Process steam consumption (300 F saturated)	.08 lb steam/lb gas
	Pressure drop	50 psia
	Fuel gas heating value (HHV)	121 Btu/SCF

Gas Analysis: Component	Gasifier Exit (mole fraction)	Cleanup System Exit (mole fraction)
H_2	.1287	.1426
CO	.2066	.2289
CH_4	.0009	.0010
N_2	.5592	.6195
H_2O	.0682	.0000
CO_2	.0266	.0000
A	.0072	.0080
H_2S	.0024	.0000
COS	.0002	.0000

FIG. 37

EFFECT OF PRESSURE RATIO ON POWER COST

TURBINE INLET TEMPERATURE = 2200 F
TWO-PRESSURE STEAM SYSTEM
HIGH STEAM PRESSURE = 1250 PSIA
STACK TEMPERATURE = 300 F



The power cost presented is the estimated busbar cost minus the corresponding estimate for a pressure ratio of 16. In going from low- to high-pressure ratio, the oil consumption and gasifier output must increase, causing the gasifier cost to increase by 0.7 exponential cost factor. The effect of pressure on gasifier cost is neglected. In addition, the gas turbine and steam system costs also increase resulting in minimum power cost for pressure ratios near 16.

Boost Compressor Location - Varying the location of the boost compressor yielded interesting results. For the oil gasifier system described above, the results of varying gasifier boost compressor location are presented in Figure 38. The efficiency range for compression after cleanup results from two possible assumptions. If the same absolute pressure drops are maintained through the system the net efficiency would be 30.2 percent. If the same fractional pressure drops are maintained the resultant efficiency would be 30.8 percent. A number of factors tend to mitigate and actually reverse this effect in a coal fed system. Because of the lower air to fuel ratio with coal, there is a greater increase in mass flow rate in going from the air stream to the clean fuel stream. For the BuMines/low-temperature system, the fuel gas compressor uses about 50 percent more power than does the air boost compressor. While heat of compression appears in the product gas stream, locating the compressor in the low-temperature gas stream actually displaces one of the most desirable fuel gas regeneration locations resulting in a need to reject an additional amount of low-temperature heat. Thus, the boost compressor has been located in the air stream in all of the final coal systems. However, the possibility of a better location should be considered in oil-fired systems.

Parametric Analysis of Nitrogen Oxide Emissions From Gas Turbine Power Systems Burning Low-Btu Fuel Gas

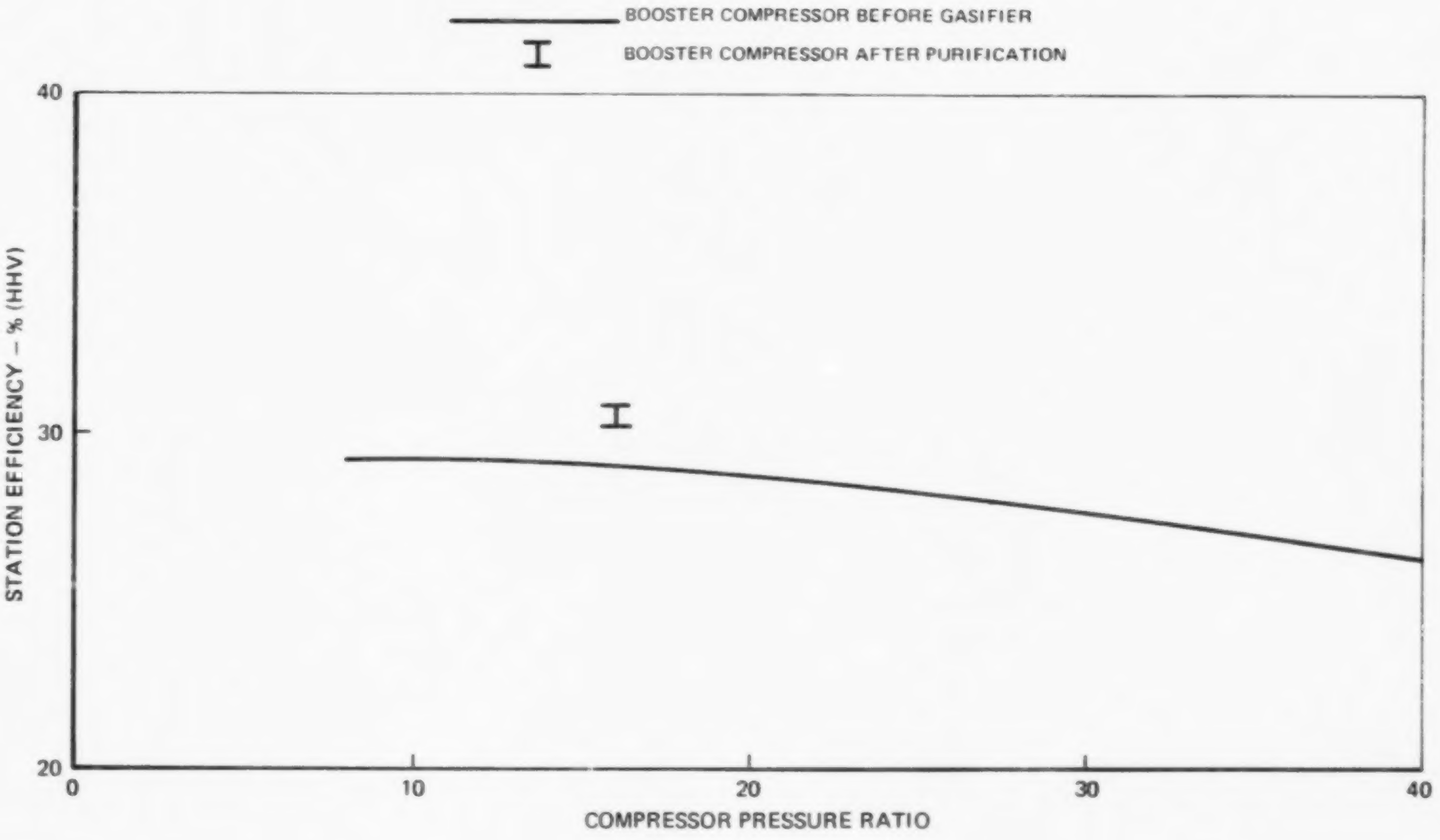
Oxides of nitrogen, commonly lumped together as NO_x are receiving increasing attention as air pollutants. The various oxides are easily interconverted in the atmosphere, their ratio depending on the action of sunlight, oxygen, and other oxidizing or reducing agents present. A major contributor of these pollutants are the nitrogen oxides formed in the hot reaction zones of all air-breathing combustion engines. They are formed primarily as NO (nitric oxide), although small quantities of NO_2 (nitrogen dioxide) and N_2O (nitrous oxide) may also be formed.

Two mechanisms are known to contribute to the formation of nitric oxide in combustion systems. The most important mechanism for gas turbines, and other systems which burn relatively clean fuels, is referred to as the thermal or hot air mechanism. In this mechanism, nitrogen and oxygen from the

FIG. 38

INTEGRATED COGAS/OIL GASIFICATION STATION PERFORMANCE

TURB. INLET TEMP. = 2200 F
TWO-PRESSURE STEAM SYSTEM
HIGH STEAM PRESSURE = 1250 PSIA
STACK TEMPERATURE = 300 F



atmosphere react in the hot combustion zone to form nitric oxide. The second mechanism is important when relatively dirty fuels such as coal and residual fuel oil are burned. Most dirty fuels contain small but significant quantities of organic nitrogen compounds Because nitrogen-carbon and nitrogen-hydrogen bond energies are so much lower than that for molecular nitrogen, much of the fuel becomes oxidized during combustion. Experimental studies(35) of the formation of nitric oxide from fuel nitrogen indicate that the formation rates are very rapid, occurring on a time scale comparable to that of the hydrocarbon combustion reactions. This mechanism is strictly fuel dependent and proceeds at lower temperatures than needed for the thermal mechanism.

Fuel nitrogen could be a problem in systems using gasified fuels. During gasification of dirty fuels, some fuel nitrogen would carry over into the raw fuel gas as combustible nitrogen compounds (primarily ammonia, with smaller concentrations of hydrogen cyanide, pyridine, pyridine bases, and acidic nitrogenous compounds). If retained in the fuel gas, these compounds could result in excessive emissions of nitrogen oxides. It appears that low-temperature cleanup systems would adequately remove undesirable nitrogen compounds, but the same cannot be said for high-temperature cleanup systems.

Status of NO_x Pollution Modeling - The chemical kinetics of NO formation via the thermal mechanism are fairly well understood(36-39). In the gas turbine, the local temperature, residence time, and species concentrations which govern NO production are controlled by engine operating conditions, the combustor internal flow field, fuel nozzle characteristics, and the air addition schedule to the burner can. Lack of an adequate analytical description of the combustor flow field and the fuel/air mixing characteristics has prevented accurate estimation of the temperature-time concentration history which is essential for reliable estimation of NO formation. At the present time, a number of engineering and research establishments, including several groups within United Technologies Corporation, are attempting to develop comprehensive gas turbine combustor models. The results of this modeling work have been very encouraging and are leading to a better understanding of NO emissions.

Despite the rather primative state of combustor/pollution models, at least one model exists which can be used to predict trends in emission characteristics under varying operating conditions for typical types of gas turbine combustors. This model (described in Appendix B) is a semiempirical simulation developed by United Technologies for use with conventional liquid fuels and subsequently modified for use with low-Btu gaseous fuels. The model uses engineering approximations of the flow field in the combustor combined with physical and chemical combustion kinetic models.

Nitric Oxide Emissions From Gas Turbine Combustors - The combustor model described in Appendix B was used to simulate the combustion of low-Btu fuel gas in a can-type combustor similar to that used in a high-performance, utility turbine (Pratt & Whitney FT4). A parametric study was made of the effect of varying operating conditions on thermal NO_x emissions for four fuel gas compositions produced by four coal gasifier-cleanup system combinations. A computer simulation was also made using test rig conditions burning gasified oil (see Table 14). These rig tests were run at atmospheric pressure with the fuel gas at ambient temperatures and are the only known rig tests of gas turbine combustors burning fuel gas from an air-blown gasifier. These rig tests were made using the Pratt & Whitney FT4 industrial gas turbine combustor can and gasified oil produced by Texaco Oil Company's pilot gasifier(40, 41). The tests went very well with no problems in burning the fuel and with low NO_x production -- so low that the NO_x measurements were only slightly higher than the error margin in the chemical measurements. The model simulation results showed good agreement with the rig tests. More, and greatly expanded, rig tests have been run in 1975, and it is anticipated that a complete correlation of the model simulation and rig test conditions can be undertaken.

Fuel gas characteristics for the two-stage gasifier being developed by Bituminous Coal Research, Inc.(6, 7) coupled with the Consolidation Coal Company half-calcined dolomite cleanup system are summarized in Table 14. When the simulated FT4 combustor (Figure 39a) is modelled under these conditions, the thermal NO_x concentration in the combustor is predicted as shown in Figure 39b in which half of a cross-section of the annular combustor is shown. An FT4 gas turbine, which develops approximately 30 Mw of electricity at peak rating, contains eight of these combustor cans. Each can contains six swirlers, shown at left in Figure 39a, symmetrically arranged about the annulus and a seventh swirler at the end of the center tube. This extends more than halfway down the combustor can and is used to inject dilution air into the main combustion area of the combustor can downstream of the recirculation zone. The recirculation zone, held by the vortical flow behind the swirlers, acts as a flame holder and ignition source for the main body of fuel in the combustor can.

While the NO_x concentration in the recirculation zone is high, because the temperatures there are high and the fuel-air ratio is near stoichiometric, the recirculation zone is not the major contributor to NO_x emissions from gas turbine combustors. As may be seen by examining Figure 39, the NO_x concentrations are higher downstream of the recirculation zone. In Figure 40 the cumulative NO_x production versus axial distance along the combustor is shown. It may be seen that virtually all of the thermal NO_x emissions are produced downstream of the recirculation zone. The small contribution of the recirculation zone to NO_x emissions is due to the relatively small flow through the

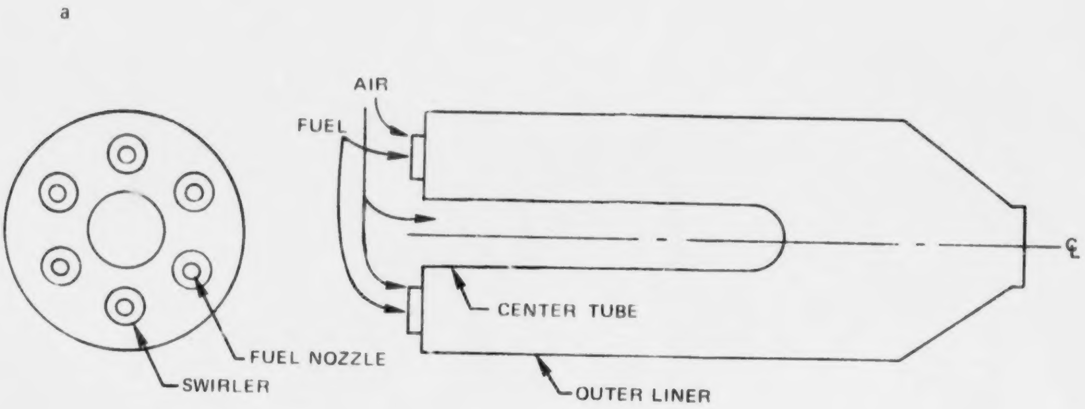
Table 14

LOW-BTU GAS CHARACTERISTICS AT DESIGN POINTS FOR FOUR GASIFIER-CLEANUP SYSTEMS

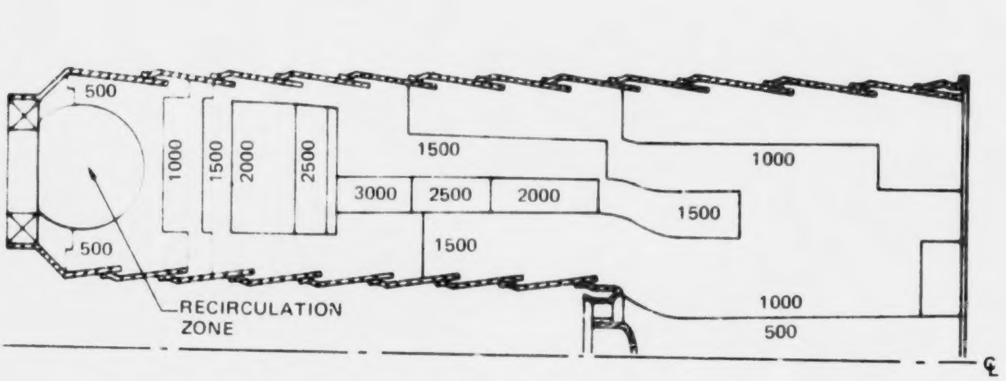
Gasifier	Cleanup	P (atm.)	T_{air} (F)	T_{fuel} (F)	H_2O	CO	H_2	CO_2	N_2	Ar	CH_4	NH_3	COS	H_2S	Ash
Gasified Coal															
BOM	Selexol	15.8	754.6	265	.0001	.2387	.1598	.0502	.5197	----	.0311	.0003	.0001	----	----
BOM	Iron Oxide	15.8	754.6	1070	.0391	.1534	.1819	.1113	.4770	----	.0277	.0063	----	.0001	.0032
BCR	Selexol	23.8	903.1	1000	.0001	.2123	.1487	.0649	.5318	----	.0421	----	.0001	----	----
BCR	Consol	23.8	903.1	1700	.0994	.1737	.1347	.0929	.4588	----	.0356	.0043	.0001	.0005	----
Gasified Oil															
Texaco	Rectisol	1.3	732	95	----	.2218	.1579	.0001	.6108	.0079	.0015	----	----	----	----

FIG. 39

NO$_X$ IN GAS TURBINE COMBUSTORS


a
AIR
FUEL
CENTER TUBE
FUEL NOZZLE
SWIRLER
OUTER LINER
℄

FT4 COMBUSTOR CONFIGURATION

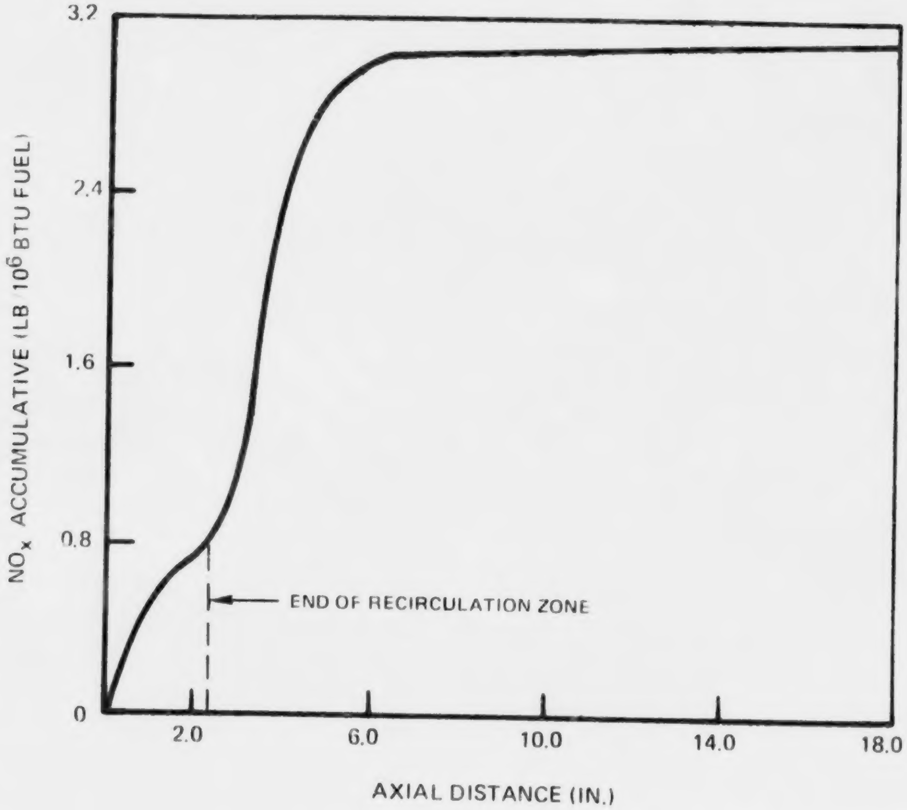

b
500
1000
1500
2000
2500
3000
2500
2000
1500
1000
1500
1500
500
RECIRCULATION ZONE
1000
500
℄

(PPM BY WEIGHT)
PRATT & WHITNEY FT4 COMBUSTOR
BCR CONSOL PROCESS FUEL GAS
P - 23.8 ATM
$T_{FUEL} = 1700^\circ F$

NO$_X$ CONCENTRATION ISOLINES

FIG. 40

NO_x VS AXIAL DISTANCE



N03-80-2

recirculation zone. Since the function of the recirculation zone is to provide a heat source to stabilize the flame, only enough fuel and air are added to maintain its temperature. The main fuel combustion takes place downstream. It may be seen that the NO_x production in gas turbine combustor cans is directly related to combustion in the main burning zone downstream of the recirculation zone and, in particular, to the local temperatures since NO and NO_2 production rates are extremely dependent on temperatures.

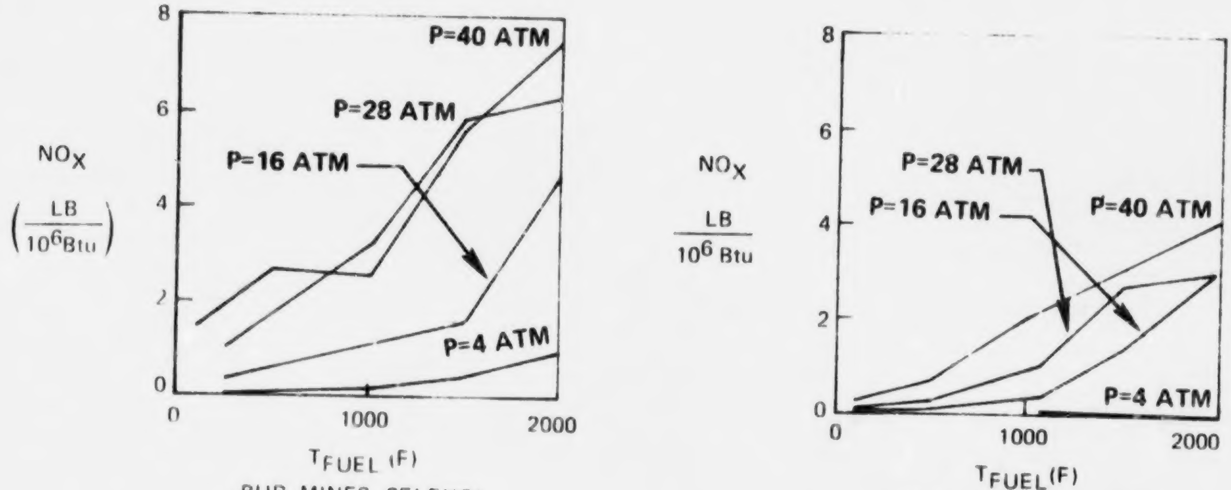
The production of thermal NO_x for a series of fuel gas temperatures are given in Figure 41 for four gasifier-cleanup systems (see Table 14). The combustion air temperature was set to correspond to the appropriate compressor discharge conditions as determined from the cycle calculations. All runs were made with air flow schedules and geometry for the Pratt & Whitney FT4 scheme 24-2DB combustor. In order to obtain a set of comparable runs, fuel flows were chosen with the objective of obtaining turbine inlet temperatures of 2200 F for the Bureau of Mines gasifier cases and 2600 F for the BCR gasifier cases. The units used in Figure 41 are lb $NO_x/10^6$ Btu where Btu refers to the higher heating value of the fuel gas. To convert to lb $NO_x/10^6$ Btu coal, it is necessary to multiply by the effective cold gas efficiencies which are 83.2 percent, 78.7 percent, 76.2 percent, and 77.1 percent for the BuMines-Selexol, BuMines-Iron Oxide, BCR-Selexol, and BCR-Consol processes respectively.

The parametric thermal NO_x estimates presented in Figure 41 reveal important trends. Increases both in fuel temperature and in cycle pressure cause dramatic increases in NO_x emissions. This comes as no surprise since both trends had been predicted some time ago(4, 42). It has previously been predicted that the temperature rise associated with one Btu of fuel gas sensible heat was twice that associated with one Btu of chemical heating value. The reason for this is simply that increased chemical heating value is caused by a higher concentration of combustibles in the fuel and, thus, requires more combustion air. Since the additional combustion air consumes most of the incremental heating value while being heated up to the initial combustion temperature, little heat remains for further temperature rise. Sensible heat, however, simply adds to the temperature rise. Since the kinetics of NO_x production is strongly temperature dependent, the trends in Figure 41 seem appropriate.

Fuel Nitrogen - As has been previously pointed out, fuel nitrogen compounds would add to the thermal NO_x estimates presented in Figure 41. For the four combinations of gasification/cleanup systems shown in Table 14, the only fuel nitrogen compound of consequence is NH_3 with concentrations ranging from 0.03 to 0.65 percent (Vol.). The mechanism whereby NH_3 is converted to NO_x is not well understood at this time, although there appears to be ample data(43-45)

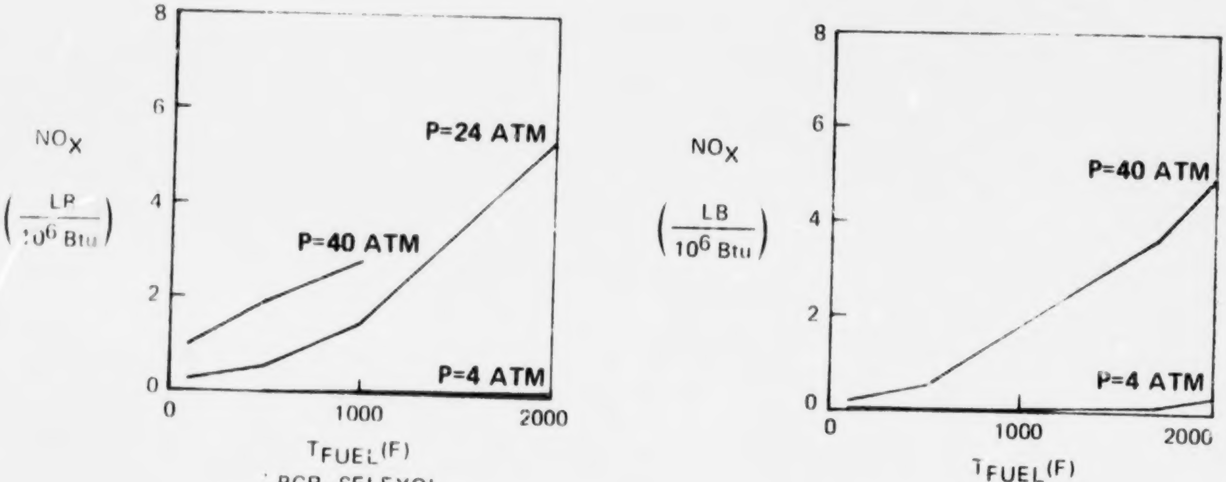
FIG. 41

NO$_X$ PRODUCTION VS. T$_{FUEL}$ AT VARIOUS PRESSURES FOR COAL–GAS FROM FOUR GASIFIER–CLEANUP SYSTEMS



BUR. MINES–SELEXOL

BUR.MINES–IRON OXIDE



BCR–SELEXOL

BCR–CONSOL

which suggest that if the concentration of NH_3 in the fuel gas is low (say < 0.5 percent, vol.), all of it will be converted to NO_x; whereas if the concentration is high, only a fraction will be converted to NO_x. While it is difficult to assign a specific fraction of conversion, the mechanism is such that the addition of NH_3, for example, to a low concentration assumed to have essentially 100 percent conversion to NO_x, will not result in a net decrease in NO_x production although less than 100 percent of the total fuel-bound nitrogen compounds are now converted to NO_x.

One of the more plausible attempts to correlate fuel nitrogen conversion data is presented by Fenimore(44). A mechanism is postulated whereby all the fuel nitrogen reacts through an intermediate species I according to the reaction:

$$I + R \rightarrow NO + \ldots\ldots\ldots \qquad (33)$$

$$I + NO \rightarrow N2 + \ldots\ldots\ldots \qquad (34)$$

where R is a nitrogen-free species which causes the eventual oxidation of I to NO. The second reaction denotes the decomposition of NO, which could become important when the NO concentration becomes high. The postulate further suggests that R might correspond to the OH radical and that I might correspond to NH_2.

In order to utilize the above model, it would be necessary to combine it with a thermal and combustor NO_x model so that local concentrations of the reactive species could be used. No meaningful attempt has yet been made to do this. Although using average species concentration can be misleading, this was done using the correlations in reference 44 to obtain an order of magnitude estimate of the NO_x conversion yield for fuel nitrogen. The effective yield was nearly 100 percent from the two cases with low NH_3 concentration (BuMines/Selexol and BCR/Selexol). For the BuMines/hot iron oxide and BCR/CONSOL cases, the effective yield appeared to be between 50 and 100 percent. Lacking more definitive information, 90 percent effective yield will be assumed in estimating the NO_x produced by nitrogen-based fuel constituents.

Discussion of Results of NO_x Modeling - Despite the above uncertainties in thermal and fuel NO_x predictions, some generalized conclusions can be made. It appears that NO_x models predict correct trends, with the values predicted within about a factor of 2 or 3 of actual production. High-temperature sulfur removal systems which do not remove NH_3 would result in NO_x emissions in excess of the EPA standards for coal-fired powerplants of 0.7 $lb/10^6$ Btu.

Since at the time of this writing no recognized standard exists, the value for a coal-fired station is used. Whether low-temperature cleanup systems could meet the NO_x standard would depend on the use of fuel gas regenerators to raise the fuel supply temperature.

Fuel gas temperatures in excess of approximately 1000 F, and pressures over approximately 16 atm, could lead to NO_x emissions in excess of the 0.7 $lb/10^6$ Btu standard. For temperatures and pressures below these values it should be possible to design gas turbine combustors which conform to EPA standards.

It must be noted that the foregoing analysis was based upon a simulated FT4-type burner can. This combustor was designed for use with liquid fuels and high-Btu gas. Tests subsequent to the original Montebello tests (previous rig tests) indicate that various modifications to these combustors could result in lower NO_x formation. Thus, future efforts in modelling NO_x formation could result in estimates more closely resembling actual conditions.

SECTION 3

COMPARISON AND SELECTION OF CLEANUP SYSTEMS

Section 3 contains comparisons of the various low- and high-temperature cleanup systems reviewed in Section 1 and describes the criteria and methods used to select these cleanup systems for integration with the gasifier and power system.

COMPARISON OF CLEANUP SYSTEMS

The comparison of the low- and high-temperature cleanup processes identified in Section 1 includes as the primary criteria the efficiency of pollutant removal, effect on power system performance, cost considerations, and estimated time of availability (first- or second-generation) for commercial application.

Low-Temperature Systems

The majority of the approximately 40 low-temperature desulfurization processes identified in Section 1 are commercially operative and, therefore, could be applicable to both first- and second-generation integrated power systems. In selecting those likely to be most applicable in treating coal derived fuel gas, the following factors were taken into consideration:

a. Sulfur removal capabilities, not only with respect to H_2S, but also other sulfur compounds such as COS and CS_2.

b. Selective absorption of sulfur compounds over carbon dioxide. Since CO_2 need not be removed from fuel gas intended for use in advanced power cycles, absorption of CO_2 represents an increased operating load on the system.

c. Type of absorbent insofar as the treated fuel gas may be contaminated by entrained or volatilized solvent which could be detrimental to downstream system components.

d. The system's tolerance to other contaminants present in the raw fuel gas such as ammonia, cyanides, phenols, and tars.

e. Overall energy requirements and operating cost.

A review of the literature of sour gas stripping was performed and those processes which appeared to be commercially important were identified. However, since the scope of this study was limited to published information, the availability of performance data allowed only ten systems to be selected for further analysis.

Of the chemical solvent processes, the Alkazid, Benfield, and Phosphate processes were selected on the basis of their capability for partial selective H_2S absorption. The Sulfinol, Selexol, Fluor Solvent, Purisol, and Rectisol systems were included as they are commercial physical solvent processes. The Stretford and Giammarco Vetrocoke systems represented commercially operating direct conversion processes. Evaluation of these ten processes was based upon treatment of 400 MMSCFD of dry feed gas having the following composition:

	N_2	CO	H_2	CO_2	CH_4	H_2S	COS	NH_3
Volume %	45.0	20.0	19.0	11.0	4.0	0.5	0.1	0.5

This composition is typical of that expected from air gasification of coal after tar, naphtha, and particulate removal. It was assumed that the gas would be available at 300 F and 400 psig for low-temperature cleanup.

Based solely on information available in the literature, the Fluor Solvent and Purisol processes were eliminated from further consideration since they are best suited for cases where the acid gas partial pressure is high, e.g., above 75 psia.(12) The Giammarco Vetrocoke and Stretford processes were similarly judged unsuitable since their application is economically limited to sulfur recovery of about 20 tons/day.(10, 13) The remaining processes were ranked in terms of the following characteristics and corresponding weighting factors:

Sulfur removal	1.5
Energy Consumption	1.0
Investment cost	0.5
Absorbent type	0.5
Operating temperature	0.5

According to this arbitrary ranking technique, the lower the rank the more attractive the system. While it would have been more desirable to have licensor operating and cost data for a given degree of sulfur removal as a means of comparison, the ranking procedure employed gives a qualitative indication of the relative attractiveness for these cleanup systems.

The degree of sulfur removal, both with respect to H_2S and COS, was given primary importance for pollution considerations. Given secondary consideration were the steam and power requirements since these would directly relate to the overall efficiency of the integrated power system. Absorbent or solvent type was a factor insofar as it could effect downstream units, e.g., carry over of alkali salts in the treated fuel gas is detrimental to high-temperature gas turbine blade life and, therefore, protective systems for assuring complete removal would have to be incorporated.

Based on the ranking of the six processes shown in Table 15, it appeared that the Benfield chemical solvent system and the Selexol and Rectisol physical solvent systems were comparable. These were chosen for evaluation in integrated system performance. For this purpose, preliminary evaluation data were obtained from the process licensors.

High-Temperature Systems

In evaluating the relative merits of the six high-temperature processes identified in Phase I, the following factors were considered:

a. Operating temperature

b. Capability for removing sulfur compounds, COS as well as H_2S.

c. The form in which the sulfur is regenerated, H_2S, SO_2, or elemental sulfur. Elemental sulfur is the preferred form since it can be stockpiled without presenting significant pollution problems.

d. Ability to regenerate the absorbent without substantial loss in activity.

e. Overall energy requirements and operating costs.

The high temperature processes were compared relative to desulfurizing the following off-gas from a coal gasification unit operating at 1800 degrees F and 450 psia:

	N_2	H_2	CO	CO_2	CH_4	H_2O	H_2S	COS	NH_3
Volume %	45.5	13.1	18.1	8.3	3.6	10.4	0.5	0.1	0.4

Table 15

COMPARISON OF SELECTED LOW-TEMPERATURE CLEANUP PROCESSES

PROCESS	ABSORBENT TYPE	OPERATING TEMPERATURE °F	SULFUR CONCENTRATION PPM	ENERGY REQUIRED MMBTU/HR	INVESTMENT COST MM $	BANK
Benfield	Catalyzed K_2CO_3 Solution	230	100	230	3	7
Selexol	Polyethylene Glycol Ether	40	100	90	8	6
Rectisol	Methanol	-40	10	53	7	6
Sulfinol	Sulfolane + Di-Isopropanolamine	110	100-200	140	3-5	8
TPP	Aqueous Tri-Potassium Phosphate	100	300	270	3-5	11
Alkazid	Potassium Dimethyl Amino Acetate	80	300	150	3-5	10

The thermodynamic equilibrium absorption of H_2S and COS were calculated using the equilibrium constants developed during Phase I and assuming that CO shift equilibrium was attained at the absorption conditions. Table 16 summarizes the residual sulfur levels theoretically obtainable.

For operation above 1600 F, which would be compatible with a second-generation, entrained-flow gasifier, both the Consol and Air Products dolomite-base processes appear attractive. The latter is capable of reducing the sulfur level to around 100 ppm compared to 400 ppm for the Consol process but requires additional fuel for calcination. The Battelle molten-salt process also operates in this temperature range but its ability to remove sulfur compounds to acceptable levels is questionable, particularly in the range of 30-40 atm total pressure. At low salt loading, below 25 percent, residual H_2S can be reduced to 150-1000 ppm depending on the equilibrium constant, which remains to be defined experimentally. Moreover, there is no available data on COS removal capability for this process and there is the potentially serious problem of alkali metal volatility at the high operating temperature.

The Bureau of Mines iron oxide process appears most suitable for sulfur removal at temperatures below 1500 F, preferably around 1000 F. This is the operating range for first-generation fixed-bed gasifiers for which this process was originally developed. Off-gas from a high-temperature, second-generation gasifier would require cooling to the operating temperature of iron oxide and would result in a lower thermal efficiency than for integrated systems using the dolomite-based processes. An inherent disadvantage of the iron oxide process is that, upon regeneration, the sulfur is released as sulfur dioxide. This can readily be converted to sulfuric acid for sulfur recovery purposes. However, conversion to the preferred elemental sulfur form would necessitate additional processing steps to reduce part of the SO_2 to H_2S.

Since there are no commercial high-temperature desulfurization systems and hence no available operating data, a preliminary comparison of their energy requirements and operating costs was not possible. The economics of the high-temperature systems could be significantly influenced by replacement cost for the acceptor, resulting from loss of activity over repeated regeneration cycles. Acceptor life for the dolomite system has yet to be satisfactorily demonstrated. In this respect, the iron-oxide process has an advantage.

Based on the above qualitative comparison of the high-temperature processes identified in Phase I, the Bureau of Mines process appears well suited for first-generation application. Second-generation systems could equally employ the Consol or Air Products dolomite processes.

Table 15

COMPARISON OF HIGH-TEMPERATURE CLEANUP PROCESSES

PROCESS	ABSORPTION CONDITIONS		RESIDUAL SULFUR, PPM		REGENERATION CONDITIONS		SULFUR RECOVERY
	AGENT	F	H_2S	COS	AGENT	F	
Consol	Ca CO_3·Mg O	1750	340	50	Steam + CO_2	1200	H_2S
Air Products	Ca O·Mg O	1650	100	10	Steam + CO_2 Air + Fuel	1200 1900	H_2S
Bureau of Mines	Fe_2O_3 + Fly Ash	1000 1500	130 1300	20 200	Air	1500	SO_2
Battelle Northwest	Molten $CaCO_3$	1600	150-1000	?	Steam + CO_2	1100	H_2S
ICT-Meissner	Proprietary	800	(100)*	?	?	?	Sulfur

*Estimated from Literature(27)

Preliminary Performance of Integrated Cleanup Systems

Preliminary performance estimates of the overall, integrated power systems were used as a basis of comparison of selected high- and low-temperature cleanup processes. Mass and energy balances together with utility requirements were developed for the individual desulfurization processes for integrated front end systems producing clean fuel gas from coal. For the purpose of this comparison, the entrained-flow BCR gasifier was used under the following operating conditions:

Coal Type

Illinois No. 6	
Feed, lb/hr	2000

Gasifier Operation

Temperature, F	1800
Pressure, psia	500
Air, lb/lb, coal	3.422
Steam, lb/lb coal	0.567

Gasifier Production

Net Gas, SCF/lb coal	74.53
Slag, lb/lb coal	0.087

Raw Gas Analysis, Volume %

N_2	CO	CO_2	H_2	CH_4	H_2S	COS	NH_3	H_2O
47.70	16.74	8.84	11.98	3.14	0.46	0.10	0.38	10.66

The gasifier heat and material balance for these conditions is given in Table 17. There is a net heat rejection of 0.384 MMBtu/hr, available for high-pressure steam generation, resulting from cooling the recycle char stream. Raw producer gas is then processed through the gas purification section in which sulfur is recovered in its elemental form.

Integrated Low-Temperature Cleanup Systems - Based on the preliminary evaluation of low-temperature desulfurization processes, the relative performances of the Benfield, Selexol, and Rectisol processes were analyzed. Preliminary operating data obtained from process licensors (46-48) formed the basis for estimating the cleanup system performance. In each case, the acid gas removal

Table 17

MATERIAL BALANCE FOR BCR GASIFIER
(see Figure 42)

STREAM	M.W.	1 LB/HR	2 MOL/HR	3 MOL/HR	4 LB/HR	5 LB/HR	6 MOL/HR
O_2	32.00			49.60			
N_2	28.02		16.82	186.60			203.45
CO	28.01		6.65				71.38
CO_2	44.01		2.64				37.69
H_2	2.016		4.87				51.08
CH_4	16.04		1.34				13.39
H_2S	34.08						1.96
COS	60.07						0.43
NH_3	17.03						1.62
H_2O	18.02		1.88		1134.0		45.45
ASH						174.0	
TOTAL		2000.0	34.20	236.20	1134.0	174.0	426.45
F		100	300	800	578	2800	1750
PSIA			550	550	1300		450
M.W.			24.53	28.98			24.96

unit was designed to treat the raw gas to a residual sulfur content of 100 ppm. The acid gas from the regenerator was designed for a high (> 10 percent) H_2S concentration enabling the recovery of elemental sulfur via a conventional Claus plant.

A block flow diagram for the integrated cleanup systems is shown in Figure 42. Raw producer gas from the BCR gasification system at 1750 F is first cooled in the heat recovery section to about 300 F. The heat extracted is available for regenerative heating of the clean fuel gas, boiler feed water preheating, and steam generation. This section was optimized to give the highest overall COGAS plant efficiency for each case. The gas is further cooled below the dew point to about 120 F via direct water scrubbing which also removes at least 85 percent of the ammonia present in the raw gas. Sour water from this scrubbing operation is first steam stripped and the stripped gases are sent to an ammonia recovery section. The cooled, ammonia free, producer gas is desulfurized in the low-temperature acid gas removal section producing a clean fuel gas and a regenerated acid gas stream containing 13-22 percent hydrogen sulfide. Sulfur is recovered from the latter stream together with the ammonia recovery off-gas stream in a vapor phase Claus unit, complete with a tailgas treating section for recovering a minimum of 99 percent sulfur.

Material balances for the Selexol, Benfield, and Rectisol acid gas removal processes are given in Tables 18-20 respectively. Performance of these systems in producing a clean, low Btu fuel gas are summarized in Table 21 together with the utilities requirements.

As expected for a chemical solvent system, the Benfield process requires 2.5 to 3 times more low-pressure steam for solvent regeneration than do the two physical solvent systems. This is partially offset by the higher power consumption required by the Selexol and Rectisol processes for mechanical refrigeration to obtain their subambient operating temperatures.

Using these data, a preliminary analysis of the integrated power stations gave the following relative performance characteristics:

INTEGRATED LOW–TEMPERATURE CLEANUP SYSTEM

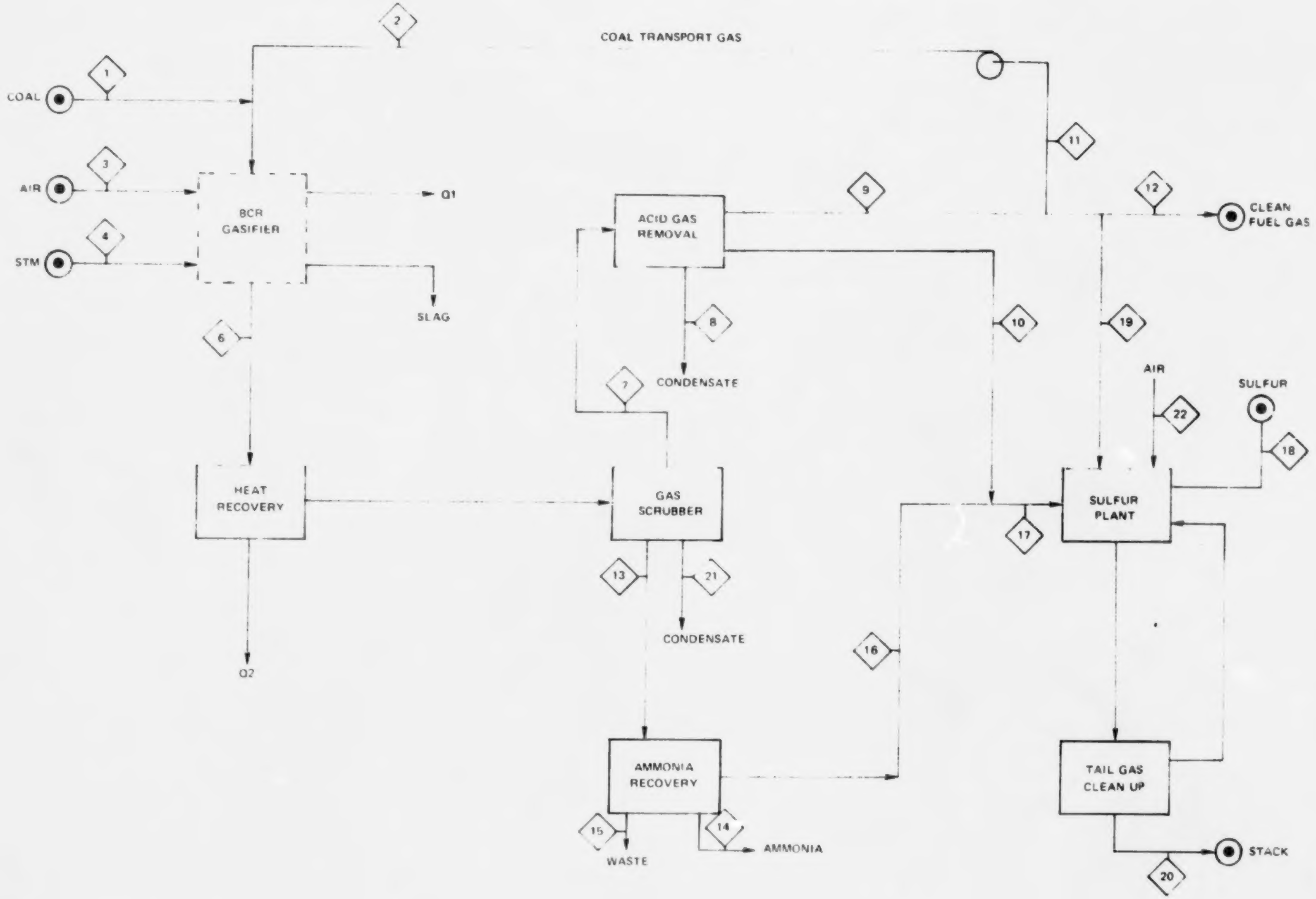

COAL TRANSPORT GAS
COAL
AIR
STM
1
2
3
4
BCR GASIFIER
Q1
SLAG
6
HEAT RECOVERY
Q2
ACID GAS REMOVAL
9
8
CONDENSATE
7
GAS SCRUBBER
13
21
CONDENSATE
AMMONIA RECOVERY
15
WASTE
14
AMMONIA
16
10
11
12
CLEAN FUEL GAS
19
17
AIR
22
SULFUR
18
SULFUR PLANT
TAIL GAS CLEAN UP
20
STACK

FIG. 42

RI 19 4

Table 18

MATERIAL BALANCE FOR SELEXOL CLEANUP SYSTEM

(see Figure 42)

COMPONENT	M.W.	STREAM 7	STREAM 8	STREAM 9	STREAM 10	STREAM 11	STREAM 12	STREAM 13	STREAM 14
O_2	32.00								
N_2	28.02	203.45		203.45		18.72	182.07		
CO	28.01	71.38		71.38		6.56	63.89		
CO_2	44.01	37.12		32.22	4.90	2.96	28.84	0.57	
H_2	2.016	51.08		51.05	0.03	4.69	45.69		
CH_4	16.04	13.39		13.35	0.04	1.23	11.95		
H_2S	34.08	1.71			1.71			0.25	
COS	60.07	0.43		0.04	0.39		0.04		
SO_2	64.06								
NH_3	17.03	0.24						1.38	1.37
NO	30.01								
H_2O	18.02	10.11	9.28	0.40	0.43	0.04	0.36	0.31	0.07
Sulfur	32.06								
TOTAL, MOL/HR		388.91	9.28	371.89	7.50	34.20	332.84	2.51	1.44
^oF		200	120	100	120	100	100	150	110
PSIA		440		430	30	430	430	30	
M.W.		25.58		25.47	40.76	25.47	25.47	24.98	

Table 18 - Continued

MATERIAL BALANCE FOR SELEXOL CLEANUP SYSTEM

COMPONENT	M.W.	STREAM 15	STREAM 16	STREAM 17	STREAM 18	STREAM 19	STREAM 20	STREAM 21	STREAM 22
O_2	32.00						0.68		3.40
N_2	28.02					2.66	15.42		12.79
CO	28.01					0.93			
CO_2	44.01		0.57	5.47		0.42	7.43		
H_2	2.016			0.03		0.67			
CH_4	16.04			0.04		0.17			
H_2S	34.08		0.25	1.96					
COS	60.07			0.39					
SO_2	64.06								
NH_3	17.03	0.01		0.24					
NO	30.01						0.24		
H_2O	18.02	0.24		0.43		0.01	3.88	35.03	
Sulfur	32.06				2.34				
		———	———	———	———	———	———	———	———
TOTAL, MOL/HR		0.25	0.82	8.56	2.34	4.86	27.65	35.03	16.19
°F		125	125	120	300	100	130	125	100
PSIA			28	28					
M.W.			40.98	40.12		25.47	31.03		28.86

Table 19

MATERIAL BALANCE FOR BENFIELD CLEANUP SYSTEM
(see Figure 42)

COMPONENT	M.W.	STREAM 7	STREAM 8	STREAM 9	STREAM 10	STREAM 11	STREAM 12	STREAM 13	STREAM 14
O_2	32.00								
N_2	28.02	203.45		203.45		17.89	182.89		
CO	28.01	71.38		71.38		6.28	64.17		
CO_2	44.01	37.12		27.29	9.83	2.40	24.53	0.57	
H_2	2.016	51.08		51.08		4.49	45.92		
CH_4	16.04	13.39		13.39		1.18	12.04		
H_2S	34.08	1.71		0.02	1.69		0.02	0.25	
COS	60.07	0.43		0.01	0.42		0.01		
SO_2	64.06								
NH_3	17.03	0.24		0.24		0.02	0.22	1.38	1.37
NO	30.01								
H_2O	18.02	10.11	(12.85)	22.00	0.96	1.94	19.78	0.31	0.07
Sulfur	32.06								
TOTAL, MOL/HR		388.91	(12.85)	388.86	12.90	34.20	349.58	2.51	1.44
°F		200	200	250	130	250	250	150	110
PSIA		440		430	30	430	430	30	
M.W.		25.58		24.81	41.30	24.81	24.81	24.98	

Table 19 - Continued

MATERIAL BALANCE FOR BENFIELD CLEANUP SYSTEM

COMPONENT	M.W.	STREAM 15	STREAM 16	STREAM 17	STREAM 18	STREAM 19	STREAM 20	STREAM 21	STREAM 22
O_2	32.00						0.58		2.92
N_2	28.02					2.67	13.62		10.98
CO	28.01					0.94			
CO_2	44.01		0.57	10.40		0.36	12.28		
H_2	2.016					0.67			
CH_4	16.04					0.18			
H_2S	34.08		0.25	1.94					
COS	60.07			0.42					
SO_2	64.06						0.01		
NH_3	17.03	0.01							
NO	30.01								
H_2O	18.02	0.24		0.96		0.29	4.20	22.18	
Sulfur	32.06				2.34				
TOTAL, MOL/HR		0.25	0.82	13.72	2.34	5.11	30.69	22.18	13.90
^oF		125	110	130	300	250	130	125	80
PSIA			28	28					
M.W.			40.98	41.28		24.81	33.14		28.86

Table 20

MATERIAL BALANCE FOR RECTISOL CLEANUP SYSTEM

(see Figure 42)

COMPONENT	M.W.	STREAM 7	STREAM 8	STREAM 9	STREAM 10	STREAM 11	STREAM 12	STREAM 13	STREAM 14
O_2	32.00								
N_2	28.02	203.45		203.45		18.87	181.87		
CO	28.01	71.38		71.38		6.63	63.81		
CO_2	44.01	37.12		29.04	8.08	2.70	25.96	0.57	
H_2	2.016	51.08		51.08		4.75	45.65		
CH_4	16.04	13.39		13.39		1.24	11.97		
H_2S	34.08	1.71			1.71			0.25	
COS	60.07	0.43			0.43				
SO_2	64.06								
NH_3	17.03	0.24		0.24		0.02	0.21	1.38	1.37
NO	30.01								
H_2O	18.02	10.11	10.11					0.31	0.07
Sulfur	32.06								
TOTAL, MOL/HR		388.91	10.11	368.58	10.22	34.21	329.47	2.51	1.44
°F		200	90	90	120	90	90	150	110
PSIA		440		430	30	430	430	30	
M.W.		25.58		25.31	43.03	25.31	25.31	24.98	

Table 20 - Continued

MATERIAL BALANCE FOR RECTISOL CLEANUP SYSTEM

COMPONENT	M.W.	STREAM 15	STREAM 16	STREAM 17	STREAM 18	STREAM 19	STREAM 20	STREAM 21	STREAM 22
O_2	32.00						0.59		3.40
N_2	28.02					2.71	13.78		12.79
CO	28.01					0.95			
CO_2	44.01		0.57	8.65		0.39	10.79		
H_2	2.016					0.68			
CH_4	16.04					0.18			
H_2S	34.08		0.25	1.96					
COS	60.07			0.43					
SO_2	64.06						0.01		
NH_3	17.03	0.01							
NO	30.01								
H_2O	18.02	0.24					2.63	35.03	
Sulfur	32.06				2.38				
TOTAL, MOL/HR		0.25	0.82	11.04	2.38	4.91	27.60	35.03	16.19
°F		125	110	120	300	90	130	125	30
PSIA			28	28		430			
M.W.			40.98	42.88		25.31	33.29		28.86

Table 21

SUMMARY OF LOW-TEMPERATURE INTEGRATED SYSTEMS

Process		Selexol	Benfield	Rectisol
Feed Streams(1)				
BCR Gas				
Flow, mph		426.45	426.45	426.45
T	F	1750	1750	1750
P	psia	450	450	450
Product Stream				
Sulfur				
Flow. lb/hr		74.9	74.9	76.2
T	F	300	300	300
P	psia	-	-	-
Transport Gas				
Flow, mols/hr		34.20	34.20	34.21
T	F	300	300	300
P	psia	550	550	550
Product Gas				
Flow, mols/hr		332.84	349.47	329.46
T	F	100	250	90
P	psia	430	430	430
Vol%	N_2	54.70	52.32	55.20
	CO	19.20	18.36	19.37
	CO_2	8.66	7.02	7.88
	H_2	13.73	13.14	13 86
	CH_4	3.59	3.44	3.63
	H_2O	0.11	5.66	
	NH_3		0.06	0.06
	H_2S	100 ppm	100 ppm	10 ppm
M. W.		25.47	24.81	25.31
HHV Btu/scf		142.6	136.5	144.0
Utilities				
Cooling Duty, MMBtu/hr		2.858	3.287	3.008
Steam @ 1300 psia, lb/hr		106.5	106.5	106.5
@ 65 psia, lb/hr		1020.0	2504 7	779.6
Electric Power, kw		60.8	25.5	41.8
Boiler Feed Water, lb/hr		219.2	222.8	223.1
Steam Condensate, lb/hr		1233.3	2718.3	994.8
Feed Gas Cooling, MMBtu/hr(2)		4.862	5.156	4.616

(1) Based on 2000 lb/hr Coal Feed to Gasifier

(2) Available for STM Generation and/or Gas Reheat

Cleanup System	Clean Fuel HHV Btu/SCF	Fuel Gas Regen. Temp. F	Gasifier Cold Gas Efficiency % Coal HHV	Overall Station Efficiency % Coal HHV
Selexol	142.6	685	73.4	31.2
Benfield	136.5	685	74.9	30.5
Rectisol	144.0	685	73.4	31.4

The above efficiencies do not necessarily represent the optimum power cycle configuration and fuel gas regenerative heating that can be achieved. However, in terms of their relative magnitudes for a particular system configuration, it can be concluded that the three low-temperature desulfurization processes can give comparable performance.

Integrated High-Temperature Cleanup Systems - Evaluation of integrated high-temperature cleanup systems included four desulfurization processes which appeared to have potential for the 1980 decade. These were the Consol and Air Products dolomite systems, the Bureau of Mines iron oxide process, and the Battelle molten salt process. Schematic flowsheets and corresponding material balances are shown in Figures 43-46 and Tables 22-25, respectively.

The flow scheme for the half-calcined dolomite process is basically that which is currently proposed by Consolidation Coal Company.(20) Both the absorption and regeneration of H_2S is conducted in fluidized beds and recovery of elemental sulfur is via the liquid phase Wackenroeder reaction. Characteristic absorption of H_2S from the BCR gasifier raw gas is illustrated in Figure 47. The partial pressure of CO_2 in the product gas is shown as a solid line for each of two different operating pressures (the variation with temperature is due to the water gas shift reaction). The limiting or minimum partial pressure shown by the dashed line is the equilibrium partial pressure of CO_2 above $CaCO_3$. The partial pressure of CO_2 must be greater than this to prevent decomposition of the $CaCO_3$ to CaO. The intercept (Point A) indicates that at the absorber operating pressure of 450 psia, the maximum operating temperature could be 1780 F and the residual sulfur would be 325 ppm (Point A'). However, the product gas enters the absorber at only 1750 F and heat balance constraints limit the maximum temperature to about 1650 F since the combined chemical reactions are endothermic to the extent of 112 Btu/mol of feed gas. The residual sulfur content at this temperature is around 600 ppm (Point B in Figure 47).

In the absorption step, 100 percent approach to equilibrium was assumed for the H_2S and COS absorption reactions and for the CO shift reaction. Regeneration of the dolomite acceptor occurs at 1300 F with an 85 percent approach to H_2S equilibrium and a 2:1 molar ratio of CO_2 to steam. The CO_2 required for regeneration is obtained by treating a portion of the clean fuel gas in a

INTEGRATED CONSOL HIGH–TEMPERATURE CLEANUP SYSTEM

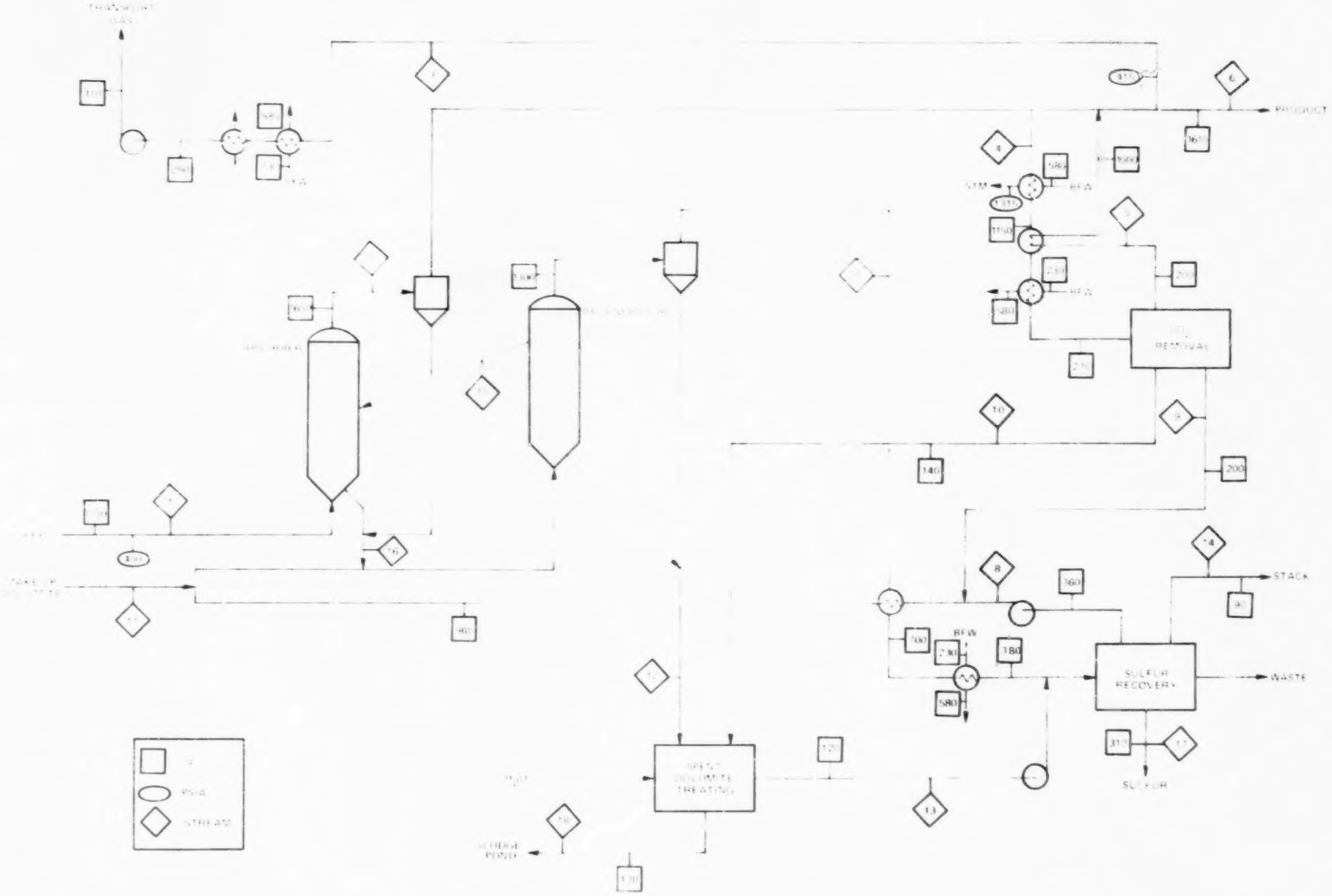

TRANSPORT GAS
PRODUCT
STM
BFW
REMOVAL
STACK
WASTE
SULFUR RECOVERY
SULFUR
SPENT DOLOMITE TREATING
SLUDGE POND
TAKE UP DOLOMITE
6
4
5
10
9
8
14
17
13
18
16
580
1150
230
200
210
140
360
90
700
380
310
120
130
250
PSIA
STREAM

FIG. 43

INTEGRATED CALCINED DOLOMITE HIGH–TEMPERATURE CLEANUP SYSTEM

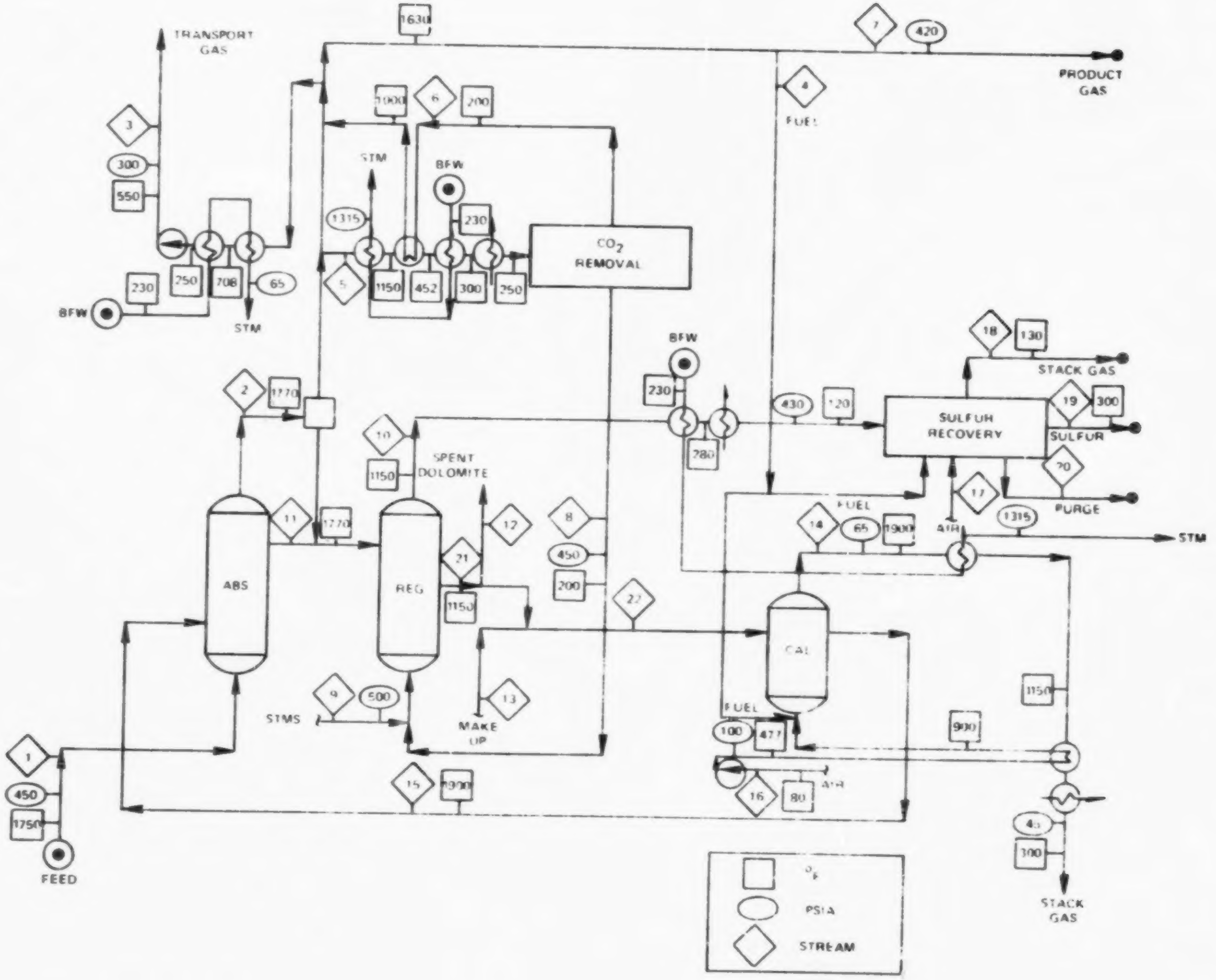

TRANSPORT GAS
3
300
550
230
BFW
250
708
65
STM
1630
1000
6
200
STM
BFW
1315
230
5
1150
452
300
250
CO_2 REMOVAL
7
420
PRODUCT GAS
4
FUEL
2
1770
11
1770
10
1150
SPENT DOLOMITE
12
21
1150
8
450
200
22
ABS
REG
STMS
9
500
13
MAKE UP
1
450
1750
FEED
15
1900
BFW
230
280
430
120
SULFUR RECOVERY
18
130
STACK GAS
19
300
SULFUR
20
PURGE
17
AIR
1315
STM
14
FUEL
65
1900
CAL
FUEL
100
477
16
80
AIR
900
1150
45
300
STACK GAS
°F
PSIA
STREAM

FIG. 44

R1-19-6

INTEGRATED BUREAU OF MINES HIGH–TEMPERATURE CLEANUP SYSTEM

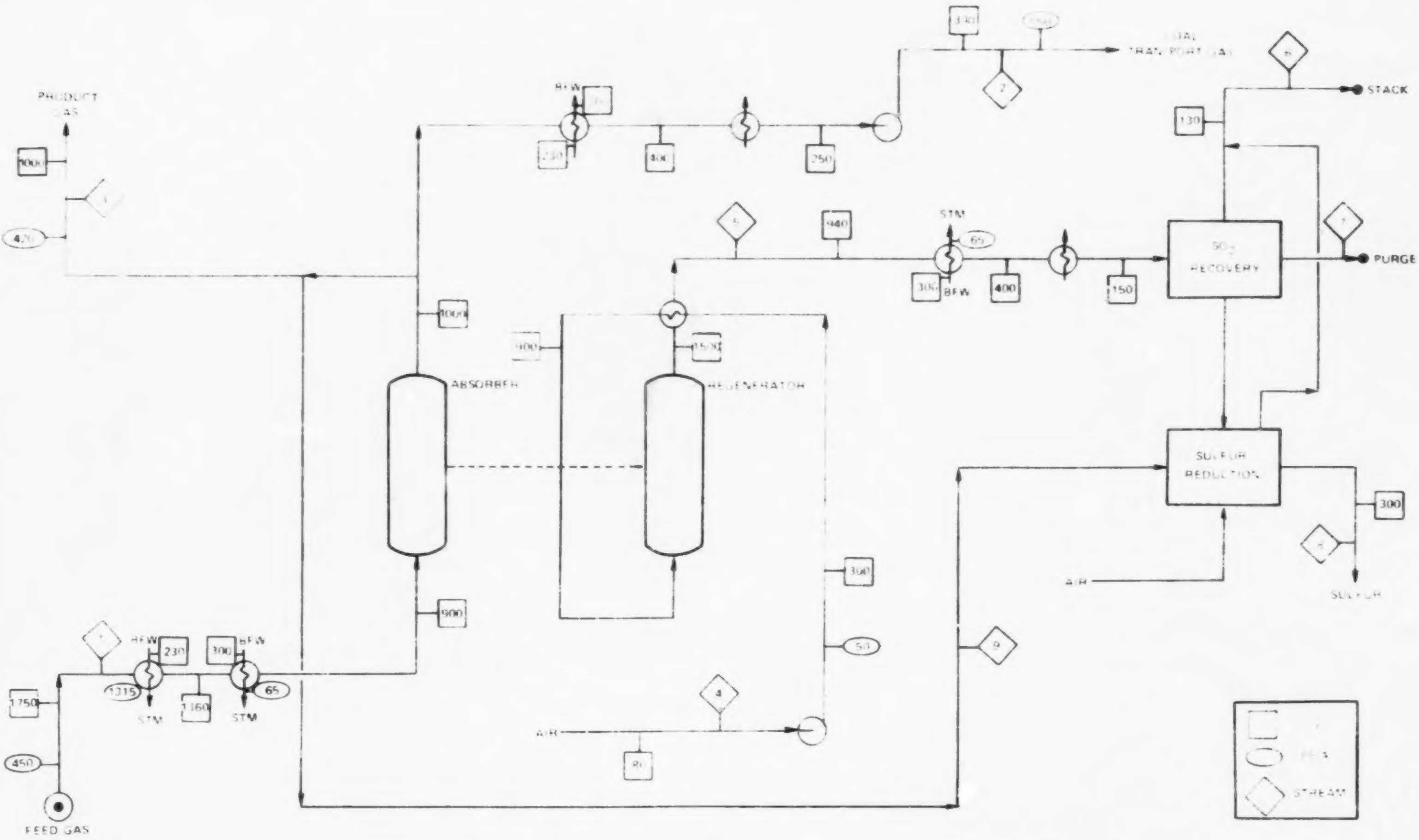
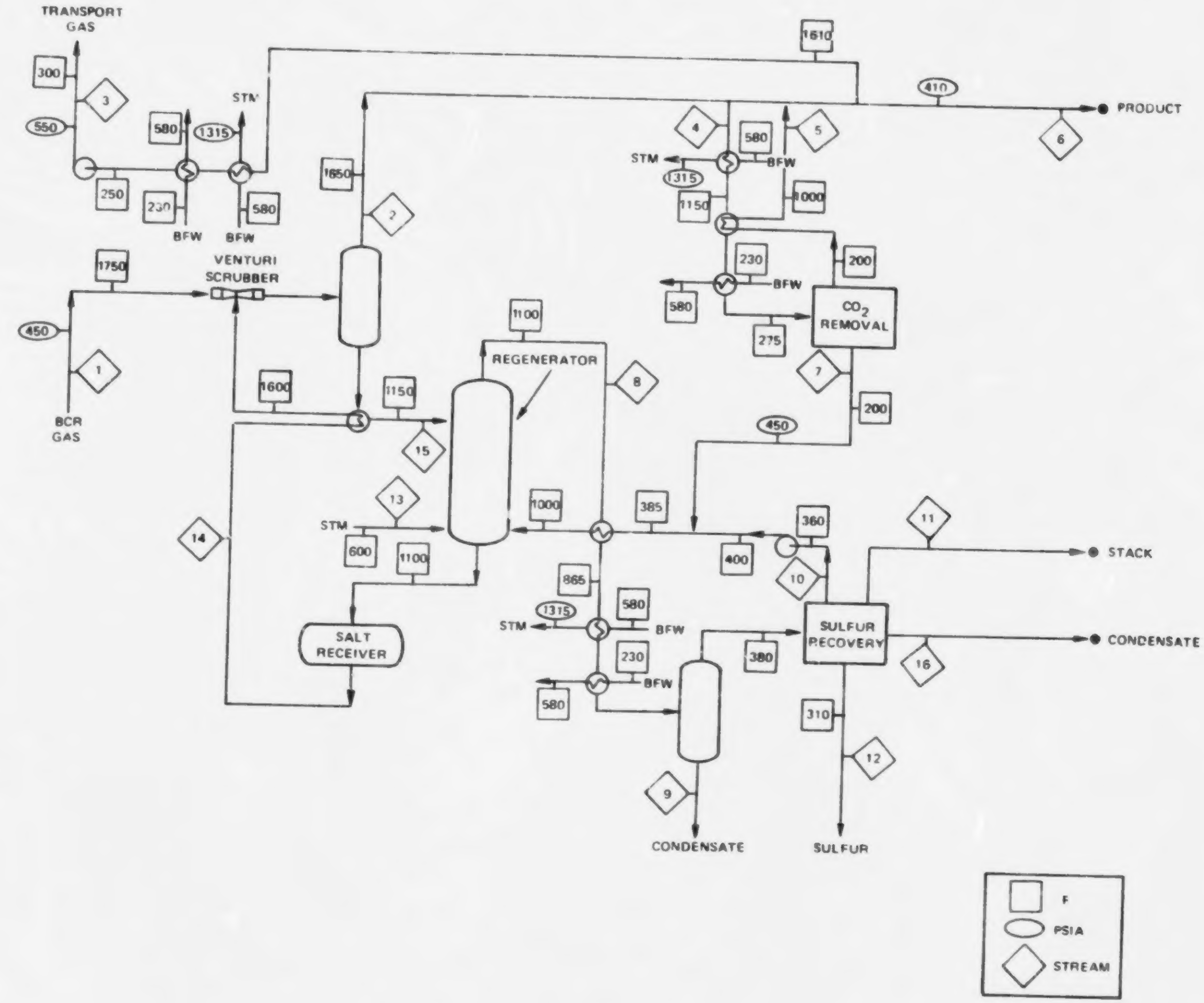

PRODUCT GAS
1000
420
BFW
230
400
250
330
2
TOTAL TRANSPORT GAS
6
STACK
130
5
940
STM
65
300
BFW
400
150
SO_2 RECOVERY
7
PURGE
1000
900
1500
ABSORBER
REGENERATOR
SULFUR REDUCTION
300
8
SULFUR
AIR
300
50
9
BFW
230
300
BFW
1750
1315
STM
1160
65
STM
4
AIR
80
450
FEED GAS
STREAM

FIG. 45

R1 19 7

FIG. 46

INTEGRATED BATTELLE MOLTEN SALT HIGH–TEMPERATURE CLEANUP SYSTEM


TRANSPORT GAS
300
550
3
250
230
580
BFW
BFW
STM
1315
580
1610
410
PRODUCT
6
1650
2
1750
VENTURI SCRUBBER
450
1
BCR GAS
1600
1150
15
4
580
5
STM
1315
BFW
1150
1000
230
BFW
200
580
275
CO₂ REMOVAL
7
200
1100
REGENERATOR
8
450
13
STM
600
1100
14
SALT RECEIVER
1000
385
360
400
10
11
STACK
865
1315
STM
580
BFW
230
BFW
580
380
SULFUR RECOVERY
CONDENSATE
16
310
12
9
CONDENSATE
SULFUR
F
PSIA
STREAM

R1-19-8

Table 22

MATERIAL BALANCE FOR CONSOL CLEANUP SYSTEM
(see Figure 43)

STREAM		1	2	3	4	5	6	7	8
	M.W.	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH
N_2	28.02	47.708	47.708	3.845	2.917	2.917	43.863		
CO	28.01	16.738	16.009	1.290	0.979	0.979	14.719		
CO_2	44.01	18.838	10.156	0.778	0.621	0.124	8.881	4.606	4.623
H_2	2.016	11.978	12.707	1.024	0.777	0.777	11.683		
CH_4	16.04	3.140	3.140	0.253	0.192	0.192	2.887		
H_2S	34.08	0.460	0.055	0.004	0.003	0.003	0.051	0.520	0.055
COS	60.08	0.100	0.008	0.001			0.007		
NH_3	17.03	0.380	0.380	0.031	0.023	0.023	0.349		
H_2O	18.02	10.658	10.334	0.793	0.632	0.143	9.052	2.303	2.744
TOTAL		100.000	100.497	8.019	6.144	5.158	91.492	7.429	7.422

STREAM		9	10	11	12	13	14	15	16
	M.W.	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH
N_2	28.02						0.939		
O_2	32.01								
CO_2	44.01	0.413	0.084			0.017			
H_2S	34.08					0.029			
SO_2	64.07						0.003		
H_2O	18.02	0.027	0.014			0.005	0.003		
S	32.07								
$CaCO_3$ $MgCO_3$	184.01			0.038					
$CaCO_3$ MgO	140.41				0.009			0.940	0.443
CaS MgO	112.46				0.029			2.883	3.380
INERTS	100			0.007	0.007			0.666	0.666
TOTAL		0.440	0.098	0.045	0.045	0.051	0.945	4.489	4.489

Table 23

MATERIAL BALANCE FOR CALCINED DOLOMITE
(see Figure 44)

STREAM		1	2	3	4	5	6	7	8
	M.W.	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH
N_2	28.02	47.708	47.708	3.884	1.355	9.161	9.161	42.469	
CO	28.01	16.738	16.376	1.333	0.465	3.145	3.145	14.578	
CO_2	44.01	8.838	9.255	0.638	0.228	1.778	0.356	6.973	1.422
H_2	2.016	11.978	12.339	1.005	0.300	2.369	2.369	10.384	
CH_4	16.04	3.140	3.140	0.256	0.089	0.603	0.603	2.795	
H_2S	34.08	0.460	0.029	0.002		0.006	0.006	0.027	
COS	60.08	0.100	0.005			0.001	0.001	0.005	
NH_3	17.03	0.380	0.380	0.031	0.011	0.073	0.073	0.338	
H_2O	18.02	10.658	10.728	0.870	0.303	2.060	2.023	9.518	0.037
TOTAL		100.000	99.960	8.019	2.759	19.196	17.737	87.687	1.459

STREAM		9	10	14	16	17	18	19	20
	M.W.	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH
O_2	32.00				0.390	0.428	0.056		
N_2	28.02			2.470	1.468	1.610	1.947		
CO_2	44.01		0.976	1.122			1.114		
SO_2	64.06						0.002		
H_2O	18.02	1.385	0.976	0.597			1.558		
H_2	2.016			0.019					
SULFUR	32.06							0.522	
CO	28.01			0.067					
H_2S	34.08		0.446						
NO	30.01						0.003		
Na_2SO_3	126.05								0.002
TOTAL		1.385	2.398	4.275	1.858	2.038	4.680	0.522	0.002

Table 23 - Continued

STREAM		11	12	13	15	21	22
	M.W.	MPH	MPH	MPH	MPH	MPH	MPH
$CaCO_3$ $MgCO_3$	184.01			0.013			0.013
$CaCO_3$ MgO	140.41	0.040	0.011			0.564	0.553
CaO MgO	112.39				0.566		
CaS MgO	112.46	0.619	0.002		0.093	0.095	0.093
Na_2SO_3	126.05						
Inerts	100	0.122	0.002	0.002	0.122	0.122	0.122
TOTAL		0.781	0.015	0.015	0.781	0.781	0.781

Table 24

MATERIAL BALANCE FOR BUREAU OF MINES CLEANUP SYSTEM
(see Figure 45)

STREAM		1	2	3	4	5	6	7	8	9
	M.W.	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH	MPH
O_2	32.00				2.945	2.121	2.659			
N_2	28.02	47.708	3.826	40.835	11.079	11.079	20.804			3.047
CO	28.01	14.738	0.976	9.993						0.746
CO_2	44.01	8.838	1.123	11.984			2.209			0.894
H_2	2.016	11.978	1.367	14.585						1.089
CH_4	16.04	5.140	0.252	2.687						0.201
H_2S	34.08	0.460	0.001	0.008						0.001
COS	60.08	0.100								
SO_2	64.06					0.550	0.002			
NH_3	17.03	0.380	0.030	0.326						0.024
H_2O	18.02	10.458	0.445	5.174			2.296			0.386
NO							0.029			
S	32.06								0.507	
Na_2SO_3	126.05							0.037		
TOTAL:		100.000	8.020	85.592	14.024	13.750	27.999	0.037	0.507	6.388

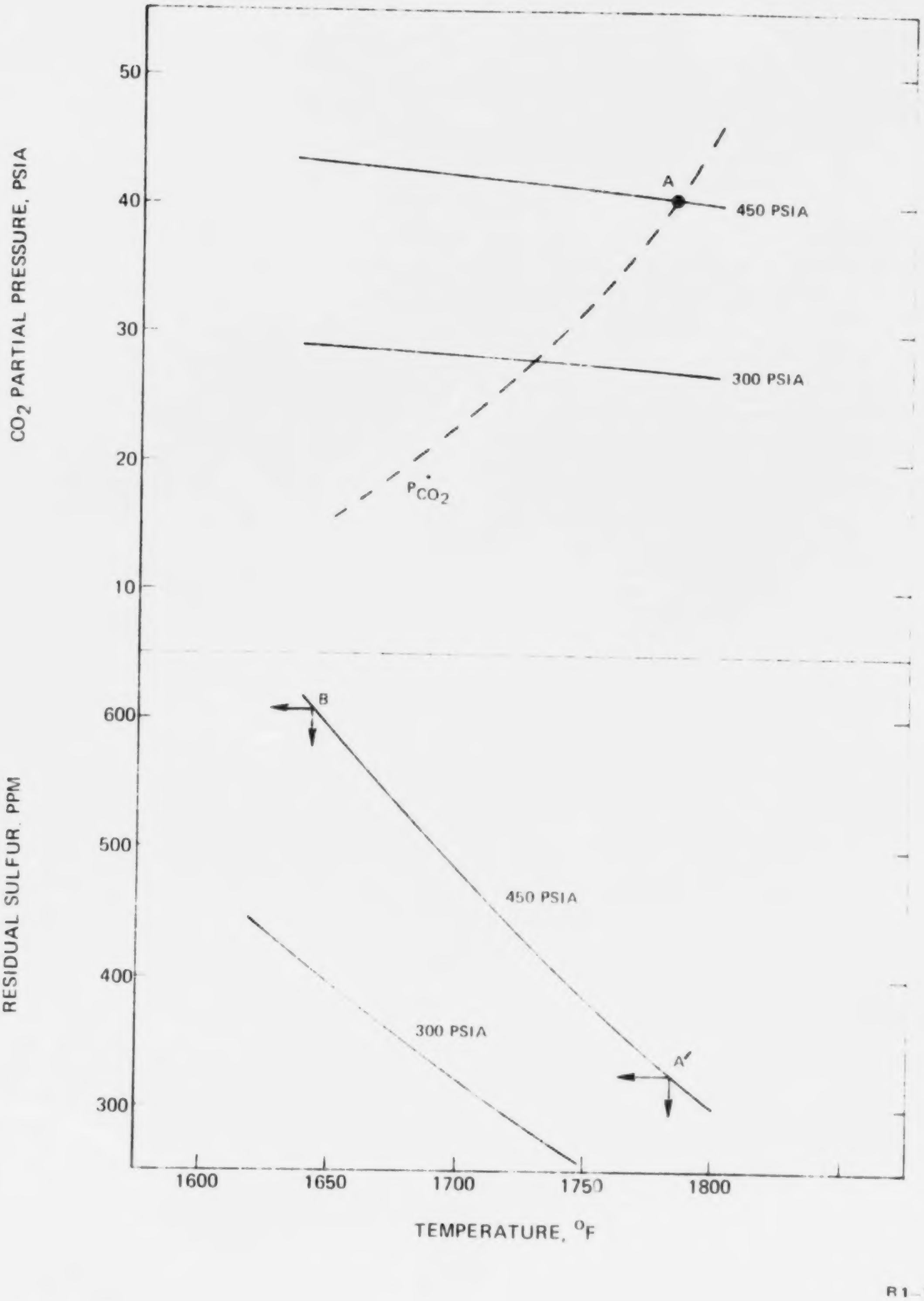
Table 25

MATERIAL BALANCE FOR BATTELLE MOLTEN SALT CLEANUP SYSTEM
(see Figure 46)

STREAM	M.W.	1 MPH	2 MPH	3 MPH	4 MPH	5 MPH	6 MPH	7 MPH	8 MPH
N_2	28.02	47.708	47.708	3.846	3.119	3.119	43.862		
CO	28.01	16.738	16.107	1.298	1.053	1.053	14.809		
CO_2	44.01	8.838	10.097	0.771	0.660	0.132	8.798	0.528	4.579
H_2	2.016	11.978	12.609	1.016	0.824	0.824	11.593		
CH_4	16.04	3.140	3.140	0.253	0.205	0.205	2.887		
H_2S	34.08	0.460	0.032	0.003	0.002	0.002	0.029		0.587
COS	60.08	0.100							
NH_3	17.03	0.380	0.380	0.031	0.025	0.025	0.349		
H_2O	18.02	10.658	10.455	0.801	0.683	0.154	9.125	0.034	18.314
TOTAL		100.000	100.528	8.019	6.571	5.514	91.452	0.562	23.480

STREAM	M.W.	9 MPH	10 MPH	11 MPH	12 MPH	13 MPH	14 MPH	15 MPH	16 MPH
N_2	28.02			1.004					
CO_2	44.01		4.579						
H_2S	34.08		0.059						
SO_2	64.07			0.003					
H_2O	18.02	11.871	2.712	0.003		16.096			3.728
Sulfur	32.07				0.525				
CaS	72.14						0.192	0.720	
$CaCO_3$	100.09						7.008	6.480	
Other	100						40.800	40.800	
TOTAL		11.871	7.350	1.010	0.525	16.096	48.000	48.000	3.728

SULFUR ABSORPTION BY HALF CALCINED DOLOMITE



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hot-potassium carbonate scrubbing unit. Spent dolomite, about 1 percent of the circulating solids, is reacted with CO_2 in a water slurry to convert the sulfide to the carbonate form before disposal to the environment In the sulfur recovery system, 90 percent of the H_2S feed is converted to elemental sulfur.

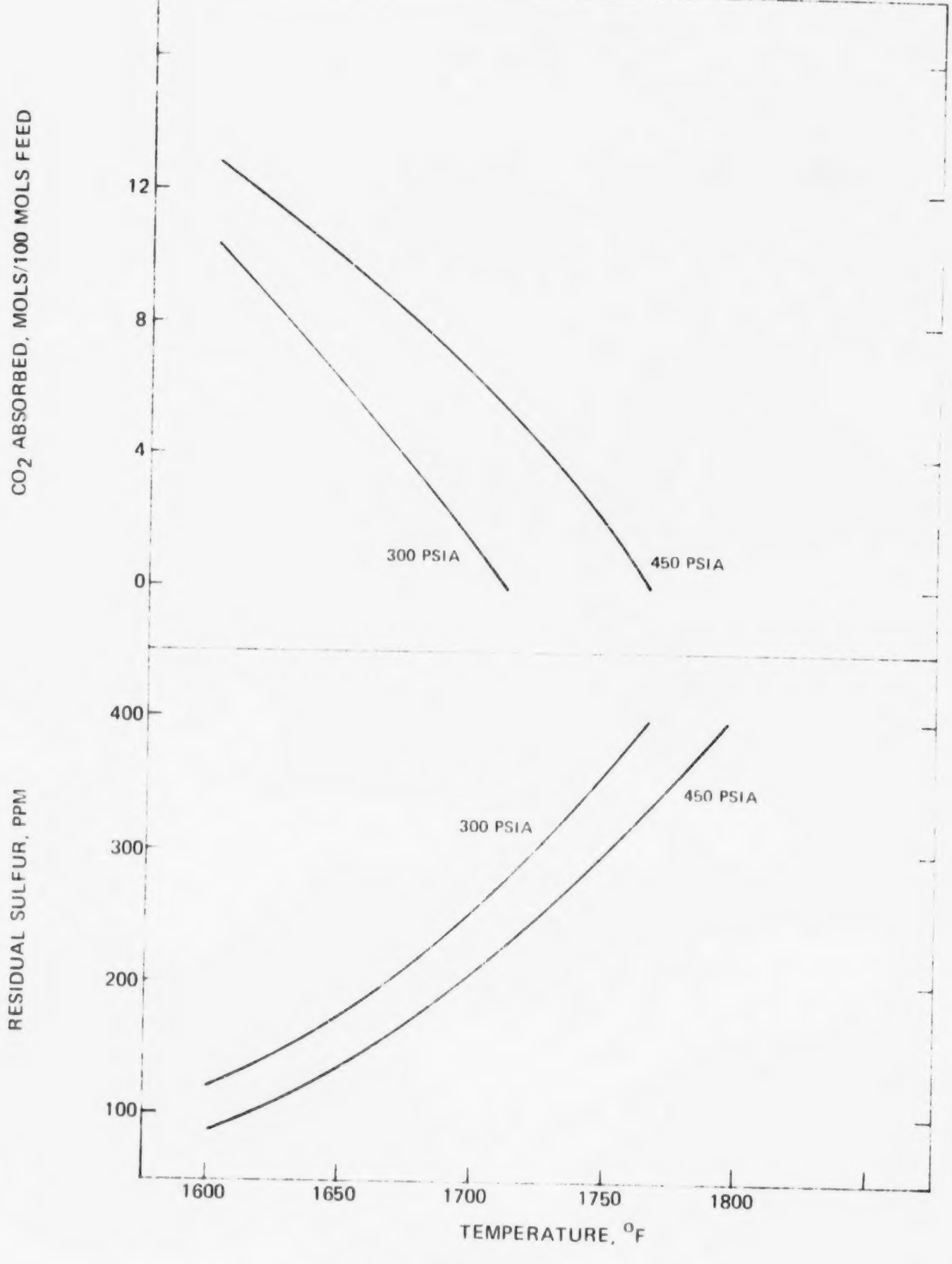
The conceptual flow scheme for the integrated Air Products calcined-dolomite system is similar to the Consol process with the addition of a calcination step. Effects of temperature on the characteristic absorption of H_2S and COS were calculated for the BCR raw gas case as shown in Figure 48. Low residual sulfur levels, on the order of 100 ppm, are obtainable by operating at temperatures of 1600-1650 F. However, at these conditions a substantial amount of CO_2 is also absorbed by the calcined dolomite acceptor. Having an exothermic heat of absorption of about 73,000 Btu/mol, absorption of CO_2 required heat removal facilities in the absorption cycle and a corresponding amount of fuel for subsequent calcination of the carbonated acceptor. The alternative is to operate at higher temperatures where the partial pressure of CO_2 is less than the equilibrium pressure over CaO so that no CO_2 absorption is possible. At this condition, above 1750 F, the lowest residual sulfur content attainable is 330 ppm.

Two case studies were therfore made for the calcined dolomite process at absorption temperatures of 1600 F and 1770 F. It was assumed that complete equilibrium was reached for sulfur and CO_2 absorption and for the CO shift reaction at the respective absorption temperature. In the low-temperature case, high-pressure steam generation is employed for removing the heat of absorption. Regeneration is accomplished with steam and CO_2 at 1150 F and, again, the required CO_2 is stripped from a portion of the product gas using a potassium carbonate scrubber. The regenerator operated at 85 percent approach to H_2S equilibrium and a 1:1 steam to CO_2 molar ratio. Under these regeneration conditions, the off-gas H_2S concentration, about 18 percent, is high enough for use as feed to a vapor phase Clause unit to recover elemental sulfur. Sulfur recovery of 99 percent is obtainable in the Claus plant equipped with a tailgas treating system. Calcination is conducted at 1900 F using combustion gases produced by burning a portion of the product fuel gas with air. The following comparison, based on 2000 lbs/hr coal feed to the gasifier, illustrates the effect of operating temperature on the calcination fuel requirements and product fuel gas characteristics:

	Case 1	Case 2
Absorption temperature, F	1600	1770
Process fuel gas, mol/hr.	95.8	11.9
Product fuel gas, mol/hr.	241.5	373.9
Product temperature F	1550	1630

FIG. 48

SULFUR ABSORPTION BY CALCINED DOLOMITE



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	Case 1	Case 2
HHV, Btu/SCF	143	127
Sulfur, ppm	105	360

High-temperature operation of the calcined dolomite system, which minimizes the process fuel requirement, would be expected to yield the more efficient performance for an integrated COGAS plant.

Based in part on process description in the literature (49), a conceptual flow scheme for the integrated Bureau of Mines iron oxide system was developed. Preliminary evaluation of this system indicated that it was most suitable for desulfurization at temperatures below 1500 F, especially around 1000 F. The case study, therefore, was based on sulfur absorption at 1000 F, and the BCR raw gas, available at 1750 F, was first cooled to operating temperature by generating high-pressure steam. Equilibrium absorption of H_2S and COS from the raw gas was calculated according the the following chemical reactions:

$$FeO + H_2S \longrightarrow FeS + H_2O \qquad (35)$$

$$FeO + COS \longrightarrow FeS + CO_2 \qquad (36)$$

With CO shift equilibrium assumed at 1000 F, the sulfur content in the treated gas is reduced to about 170 ppm. The overall absorption reactions are exothermic by 900 Btu/mol of feed.

The iron oxide system is operated in a cyclic fixed-bed fashion employing an absorbent made up of 75 percent fly ash and 25 percent Fe_2O_3. Sulfur capacity was taken at 6.0 lb/100 lb absorbent and regeneration was assumed to be 80 percent efficient. Regeneration of the sulfided absorbent is via air with a maximum temperature of 1500 F and the off-gas was assumed to have an average SO_2 concentration of 4 percent (vol). Recovery of elemental sulfur with this system is complicated by having the regenerated sulfur in the form of sulfur dioxide in low concentrations. The conceptual process, herein considered, first concentrates the SO_2 via a liquid scrubbing process on the regenerator off-gas followed by partial reduction to H_2S and then using the Claus reaction to produce elemental sulfur, i.e.

$$SO_2 + 3H_2 \longrightarrow H_2S + 2H_2O \qquad (37)$$

$$2H_2S + SO_2 \longrightarrow 3S + 2H_2O \qquad (38)$$

Reducing gas for the first reaction is supplied by a portion of the clean product fuel gas. Recovery of elemental sulfur by this process sequence is estimated at 92 percent of the SO_2 feed.

Although the preliminary evaluation of the Battelle molten salt system suggested that it was perhaps the least developed of the high-temperature processes, an estimate of its performance in an integrated system was made to assess its potential. The conceptual flow scheme was totally derived from bench scale experimental work (28). Sulfur absorption takes place in a single stage venturi scrubber where molten salt concurrently contacts the raw gas. The molten salt was assumed to contain 15 mol percent calcium salts in the form of $CaCO_3$ and CaS. Because there was no data on COS absorption, it was assumed as 100 percent and the H_2S absorption was estimated according to the reaction:

$$\mathrm{Ca\ CO_3 + H_2S \longrightarrow CaS + H_2O + CO_2} \tag{39}$$

where

$$\mathrm{H_2S} = \frac{(\mathrm{CaS})}{(\mathrm{Ca\ CO_3})} \times [\mathrm{CO_2}]\,[\mathrm{H_2O}] \times \frac{P}{K} \tag{40}$$

At 450 psia, the following H_2S absorption characteristics were calculated assuming complete CO shift equilibrium and neglecting the non-ideality of the salt solution:

	Temperature F		
	1600	1650	1700
Equilibrium Constant, K	70	90	115
H_2S residual in fuel gas, ppm			
@ 10% $CaCO_3$ conversion	450	390	270
@ 20% $CaCO_3$ conversion	1020	880	620
@ 30% $CaCO_3$ conversion	1750	1500	1060

For the purpose of this evaluation, H_2S absorption was assumed at 1650 F with 10 percent conversion of calcium carbonate resulting in a residual sulfur level in the treated gas of 390 ppm.

Regeneration of the sulfided salt melt is carried out at 1100 F with steam and carbon dioxide in the molar ration of 4:1. Salt regeneration was assumed to be 75 percent efficient with an off-gas containing 2.5 percent H_2S. As with the dolomite processes, CO_2 for regeneration is extracted from a portion of the clean fuel gas via hot potassium carbonate scrubbing. Elemental sulfur is recovered from the dilute H_2S regenerator off-gas in a liquid phase system similar to that employed by the Consol process.

A summary of the estimated operation of the four high-temperature cleanup processes is given in Table 26 for the case of treating raw gas from a BCR

Table 26

SUMMARY OF HIGH-TEMPERATURE INTEGRATED SYSTEMS

SYSTEM	CONSOL	AIR PROD. I	AIR PROD. II	BUREAU OF MINES	BATTELLE MOLTEN SALT
Feed Streams					
BCR Gas, mph					
Flow	426.45	426.45	426.45	426.45	426.45
T	1750	1750	1750	1750	1750
P	450	450	450	450	450
Product Streams					
Sulfur					
Flow	67.15	74.82	71.36	69.3	71.8
T	310	300	300	300	310
P					
Transport Gas, mph					
Flow	34.20	34.20	34.20	34.20	34.20
T	300	300	300	300	300
P	550	550	550	550	550
Product Gas					
Flow	390.17	241.47	373.94	372.85	389.99
T	1610	1550	1630	1000	1610
P	415	420	420	420	410
Vol% N_2	47.94	54.77	48.43	47.71	47.96
CO	16 09	12.16	16.63	11.68	16.19
CO_2	9.71	2.52	7.95	14 00	9.62
H_2	12.77	20.80	12.53	17.04	12 68
CH_4	3.16	3 60	3.19	3.14	3.16
H_2O	9.89	5.70	10.85	6.04	9.98
NH_3	0.38	0.44	0.39	0.38	0.38
H_2S+COS	630 ppm	105 ppm	360 ppm	170 ppm	390 ppm
LHV, Btu/scf	125.2	143.1	126.5	124.8	125.3
M.W.	24.85	21 96	24 53	24.80	24.84
Utilities					
Cooling Duty, MMBtu/hr	0.376	1.039	1.052	1.890	1.220
Steam @ 1300 psia, lb/hr	(646.9)	(8102.6)	(585.1)	(1490.0)	275.1
@ 65 psia, lb/hr					
Electric Power, MW	23.8	(322.0)	4.9	98.1	24.9
Boiler Feed Water, lb/hr	783.0	8941.4	1252.9	3745.9	1113.0
Steam Condensate, lb/hr	(122.0)	(429.9)	(348.4)	(2182.6)	129.6
Process Water, lb/hr	41.6	184.2	149.3		
Natural Gas, scf/hr					

(1) Based on 2000 lb/hr Coal Feed to Gasifier

gasifier having a coal capacity of one ton per hour. Preliminary estimates of integrated station performances are as shown below:

Cleanup System	Clean Fuel HHV Btu/SCF	Clean Fuel Temp. F	Gasifier Cold Gas Efficiency % Coal HHV	Overall Station Efficiency % Coal HHV
Consol	125.2	1610	76.1	36.0
Air Products				
Case 1	143.1	1550	53.8	29.1
Case 2	126.5	1630	73.7	35.5
Bureau of Mines	124.8	1000	72.5	31.6
Battelle Molten Salt	125.3	1610	76.1	34.9

The relative performance of the integrated systems indicates that the Consol, Air Products and molten salt systems are fairly comparable. The low relative efficiency of the Bureau of Mines integrated system is due to the low absorption temperature required for efficient desulfurization. Cooling the raw producer gas from 1750 F to 900 F, even though the heat is recovered by high-pressure steam generation, represents a loss of efficiency in terms of the relative power output from steam turbines as compared to higher efficiency gas turbines. The Bureau of Mines process is therefore more suitable for first-generation gasifiers which operate at 900-1200 F.

SELECTION OF STANDARD INTEGRATED SYSTEMS

Based on the preliminary comparison of gas purification systems, those processes which looked most attractive were selected for detailed evaluation with integrated power systems both for first- and second-generation time periods. Standard first-generation systems, judged to be commercially operative by 1980, were based on the fixed-bed gasifier concept typified by the Bureau of Mines stirred-bed gasifier. Of the low-temperature desulfurization processes currently available, the Selexol process was chosen for detailed study of the integrated system. This selection was somewhat arbitrary, since both the benfield and Rectisol processes showed comparable sulfur removal performance with Selexol in the preliminary evaluation of integrated low-temperature systems.

Although it is questionable whether any high-temperature cleanup system will be commercialized by 1980, a comparison of integrated high- and low-temperature gas purification systems was of interest in assessing the potential performance benefit of high-temperature cleanup in conjunction with first-generation gasifiers. The Bureau of Mines sintered iron oxide process was selected as the most suitable high-temperature desulfurization for this application. This selection was made on the basis of operating temperature for the first-generation gasifier, about 1000 F, and the fact that the process has been demonstrated on a pilot, stirred-bed gasifier.

Standard integrated systems for second-generation operation, during the 1980 decade, employ the two-stage, entrained-flow gasifier of the BCR type. Two standard systems were selected for detailed evaluation; one incorporating a low-temperature cleanup process and the other based on high-temperature desulfurization. For the latter, the Consol half-calcined dolomite process was selected on the basis of the preliminary comparison of integrated high-temperature cleanup systems. Although the Air Products fully calcined dolomite process appeared comparable in performance, the Consol process was judged to be further developed towards commercialization in the 1980s. Subsequent analysis has indicated that the Consol process offers slightly better overall performance when integrated into the power station (Section 4).

There do no appear to be significant developments in low-temperature desulfurization technology which would alter the types of processes available for second-generation application. Therefore, the Selexol process was again selected as being representative of the low-temperature cleanup technology which would be applicable for future integrated systems.

As part of the detailed evaluation of first- and second-generation standard systems, heat and mass balances, utilities requirements, investment cost estimates, and definition of pollutant streams were developed for the various combinations of coal gasification and gas cleanup systems selected. The evaluations were based on a coal feed rate of 8400 tons/day which roughly corresponds to a 1000-Mw COGAS power station output. The following gas turbine operating conditions, judged to be attainable for the respective time periods were specified, together with the corresponding process air inlet conditions and fuel gas delivery pressure for the gasifier/purification system plant section.

Generation	First	Second
Gas turbine inlet, F	2200	2600
Compressor pressure ratio	16:1	24:1
Process air inlet, F	800	800
Process air inlet, psia	465	600
Fuel Gas Delivery, psia	355	470

First-Generation Standard Systems

The two gasifier/cleanup system combinations considered for this time frame were:

(1) Bureau of Mines/Selexol
(2) Bureau of Mines/Sintered Iron Oxide

For each case study, the gasifier capacity was designed for 350 tons per hour of Western Kentucky No. 9 coal having the following analysis and heating value:

	C	H	S	O	N	ASH	H_2O
Wt %	64.47	5.48	3.90	4.18	1.58	15.44	4.95

HHV = 11450 Btu/lb

The gasifier operating conditions were 2.7 lb air/lb coal plus tar feed, 0.35 lb steam/lb coal plus tar, and the raw gas temperature was set at 1000 F. It was assumed that all the nitrogen contained in the coal feed was converted to ammonia, except for that present in unreacted tar. Furthermore, the raw gas was assumed to contain 0.01 percent by volume of COS in accordance with the chemical equilibrium for the hydrolysis reaction:

$$COS + H_2O \rightarrow CO_2 + H_2S \qquad (41)$$

Process flow schemes for the low-temperature and high-temperature purification cases are given in Figures 49 and 50, respectively. The corresponding process equipment lists and material balance summaries are given in Tables 27 through 30.

Bureau of Mines/Selexol System - Process Description - Coal and recycle tar are fed to the stirred bed gasification unit together with 800 F air and 1315 psia saturated steam. Under reaction conditions, analysis indicates about 98 percent of the coal carbon (including carbon in the recycled tar) is gasified. Raw producer gas leaves the gasifier at 1000 F and 420 psia and, after dust removal in the cyclone separator F-101, it is cooled to 338 F by water quench in T-201 where heavy tar components are condensed. The tar is separated from the water phase in V-201 and recycled to the gasifier. Quenched gas, saturated with water, is further cooled to 120 F by heat exchange in the sour water stripper reboiler E-402, the product gas reheater E-201, and air fan cooler E-202. Condensate consisting of water plus light oils is separated in V-202 and sent back to the quench system. The cooled gas is then water scrubbed in T-202 at 120 F to remove 90 percent of the ammonia content before being desulfurized in the Selexol unit. Bottoms from the ammonia scrubber are combined with other condensate streams from the acid gas removal system and steam stripped in T-401 where 95 percent of the ammonia and essentially all of the dissolved acid gases are removed from the water stream before discharge to the environment. The overhead gas from the stripper is cooled to 200 F and then processed in the ammonia recovery unit where 99 percent of the ammonia is recovered in the anhydrous form by extraction with an aromatic hydrocarbon solvent. Off-gas from this unit contains H_2S which is sent to the sulfur recovery system. Relatively ammonia free raw gas from T-202 is desulfurized by counter-current contact with Selexol solvent in absorber T-301. The treated gas, desulfurized to 100 ppm by volume, leaves the top of the absorber and is reheated to 265 F, via exchange with raw gas in E-201 and delivered to battery limits at 355 psia. Rich Selexol solvent, containing dissolved H_2S, CO_2, and COS, is heated against lean solvent in E-302 and then steam regenerated in T-302. Acid gas, coming overhead from the regenerator, contains about 30 percent H_2S by volume and is sent to the sulfur recovery unit.

The sulfur contained in the acid gas streams from the Selexol regenerator and ammonia recovery unit is recovered as elemental sulfur via vapor phase Claus reaction. The Claus sulfur recovery unit, shown in Figure 51, is a split flow, three stage conversion unit in which 95 percent sulfur recovery is obtained. About 1 percent of the clean product fuel gas is used to supply necessary fuel for the sulfur plant. Tailgas from the last Claus conversion stage is incinerated and discharged to the atmosphere. This results in an emission of about 0.25 lb/MMBtu of coal input. A tailgas cleanup system as the Strethford process, could reduce this at an increase in system cost.

PROCESS FLOW DIAGRAM
BUREAU OF MINES/SELEXOL SYSTEM

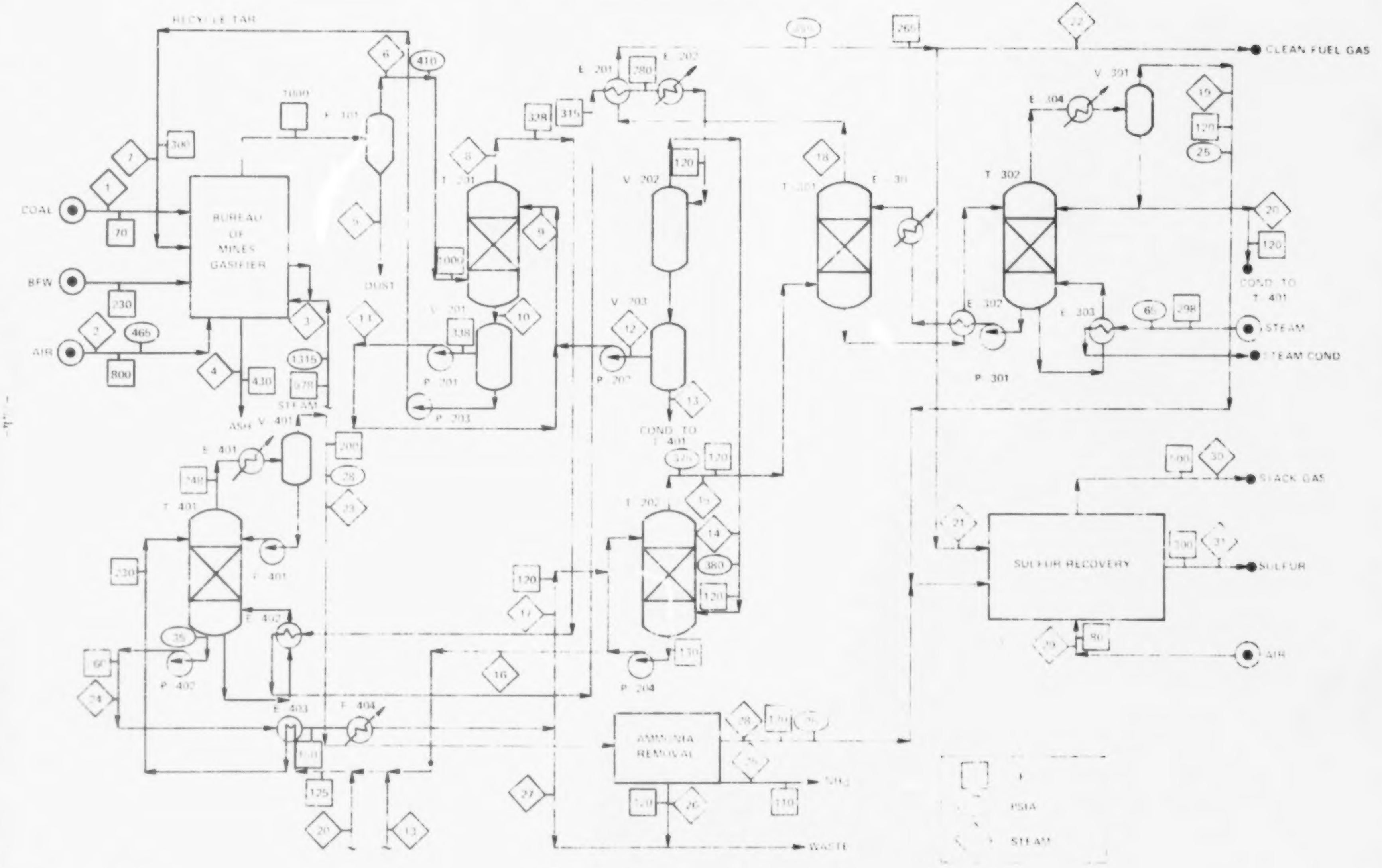
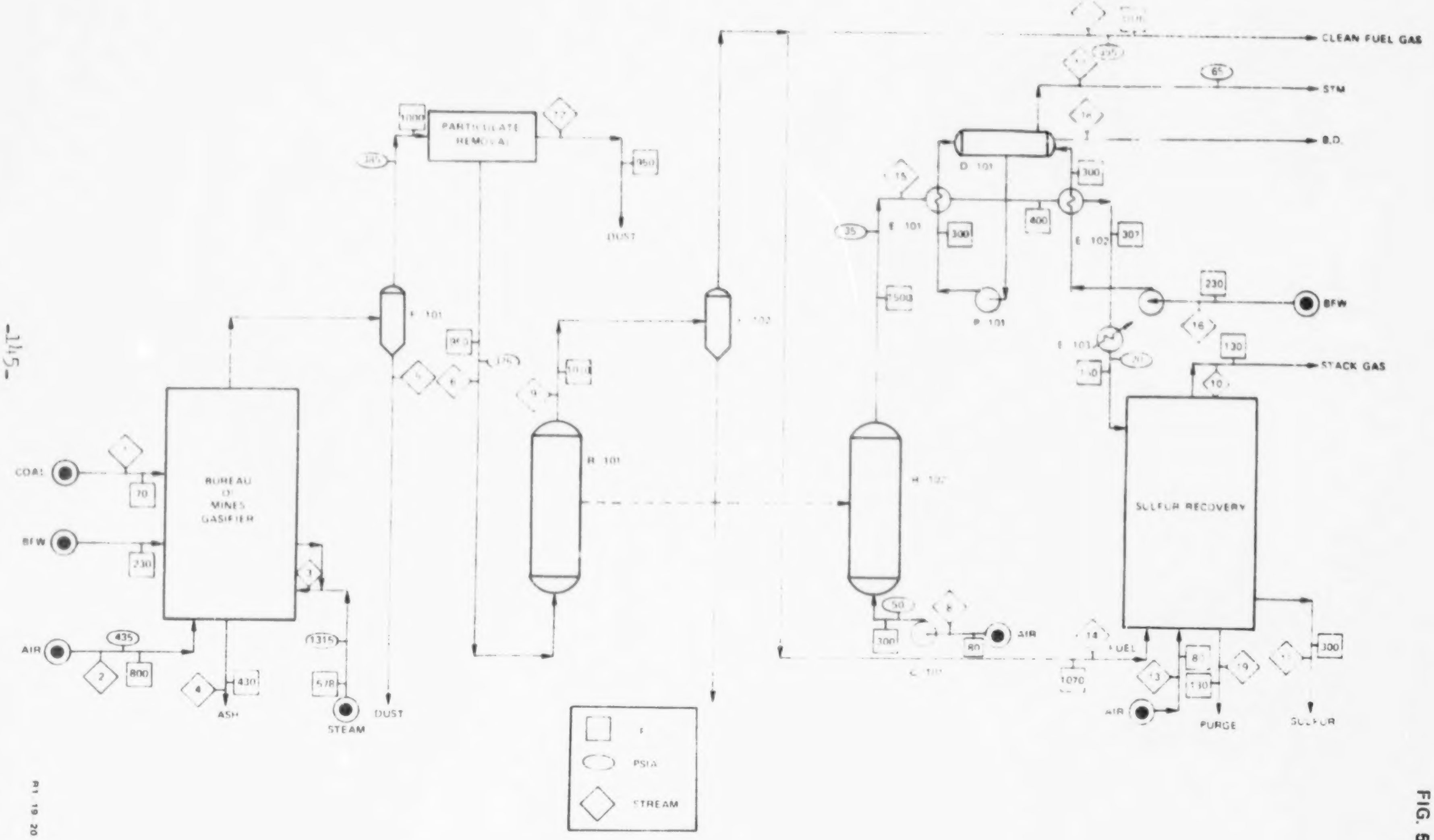

RECYCLE TAR
COAL
BFW
AIR
BUREAU OF MINES GASIFIER
ASH
STEAM
F 101
DUST
T 201
V 201
P 201
P 203
E 201
E 202
V 202
V 203
P 202
COND TO T 401
T 202
P 204
T 301
E 30
T 302
E 302
E 303
E 304
V 301
P 301
CLEAN FUEL GAS
COND TO T 401
STEAM
STEAM COND
STACK GAS
SULFUR RECOVERY
SULFUR
AIR
V 401
E 401
T 401
P 401
E 402
P 402
E 403
E 404
AMMONIA REMOVAL
NH3
WASTE
PSIA
STEAM
1
2
3
4
5
6
7
8
9
10
11
12
13
14
15
16
17
18
19
20
21
22
23
24
25
26
27
28
29
30
31
70
230
800
465
300
1000
430
1315
578
410
1000
338
328
315
280
120
326
380
130
266
265
25
65
298
500
300
90
200
28
248
220
35
60
150
125
120
110
F 401

FIG 49

PROCESS FLOW DIAGRAM
BUREAU OF MINES – SINTERED IRON OXIDE SYSTEM



CLEAN FUEL GAS
STM
65
B.D.
16
15
D 101
300
35
E 101
300
400
E 102
307
1500
P 101
230
BFW
16
E 103
130
STACK GAS
10
PARTICULATE REMOVAL
1000
345
17
950
DUST
F 101
950
5
6
375
F 102
9
1000
R 101
R 102
SULFUR RECOVERY
COAL
70
BFW
230
BUREAU OF MINES GASIFIER
3
AIR
435
2
800
4
430
ASH
1315
578
STEAM
DUST
50
8
300
AIR
80
14
FUEL
1070
C 101
13
80
130
19
AIR
PURGE
300
SULFUR
F
PSIA
STREAM

FIG. 50

R1 19 20

Table 27
BUREAU OF MINES/SELEXOL SYSTEM

EQUIPMENT LIST

SECTION 100
COAL GASIFICATION AND DUST REMOVAL

F-101 GASIFIER OFFGAS CYCLONE

SECTION 200
GAS SCRUBBING AND TAR REMOVAL

P-201 Quench Water Recycle Pump
P-202 Quench Water Pump
P-203 Tar Recycle Pump
P-204 Gas Scrubber BTMS pump
P-201 Fuel Gas Reheat Exchanger
P-202 Gas Cooler

T-201 Quench Vessel
T-202 Water Scrubber
V-201 Tar/Water Separator
V-202 Gas/Liquid Separator
V-203 Oil/Water Separator

SECTION 300
SELEXOL ACID GAS REMOVAL SYSTEM

P-301 Selexol Stripper BTMS Pump

E-301 Selexol Solvent Cooler
E-302 Rich/Lean Solvent Exchanger
E-303 Selexol Stripper Reboiler
E-304 Selexol Stripper OVHD Condenser
V-301 Selexol Stripper OVHD Accumulator
T-301 Selexol Absorber
T-302 Selexol Stripper

Table 27 - Continued

EQUIPMENT LIST

SECTION 400
SOUR WATER STRIPPING

P-401 SWS Reflux Pump
P-402 SWS BTMS Pump

E-401 SWS OVHD Condenser
E-402 SWS Reboiler
E-403 Feed/BTMs Exchanger
T-401 Sour Water Stripper
V-401 SWS OVHD Accumulator

SECTION 500
AMMONIA REMOVAL

SECTION 600
SULFUR RECOVERY

Table [illegible]

MATERIAL BALANCE FOR [illegible] MIXED/SELEXOL SYSTEM
(see Figure 4)

STREAM	M.W.	1 LB/HR	1 MOL/HR	2 LB/HR	2 MOL/HR	3 LB/HR	3 MOL/HR	4 LB/HR	4 MOL/HR
O_2	32.00			491.1[illegible]	17.[illegible]49				
N_2	28.0[illegible]			1.617.9[illegible]	57.74[illegible]				
CO	28.01								
CO_2	44.01								
H_2	2.016								
CH_4	16.04								
H_2S	34.08								
COS	60.07								
NH_3	17.03								
H_2O	18.0[illegible]					283.509	15.733		
TAR	1[illegible]								
ASH								114.132	
TOTAL		700.000		2.109.100	75.091	283.509	15.733	114.132	

STREAM	M.W.	5 LB/HR	5 MOL/HR	6 LB/HR	6 MOL/HR	7 LB/HR	7 MOL/HR	8 LB/HR	8 MOL/HR
N_2	28.0			1.617.931	57.742			1.617.931	57.742
CO	28.01			743.637	26.549			743.637	26.549
CO_2	44.01			294.691	6.696			294.691	6.696
H_2	2.016			35.772	17.744			35.772	17.744
CH_4	16.04			55.819	3.480			55.819	3.480
H_2S	34.08			27.094	795			27.094	795
COS	60.07			721	12			721	12
NH_3	17.03			13.437	789			13.437	789
H_2O	18.0			17[illegible].9[illegible]0	9.51[illegible]			813.441	45.141
TAR	1[illegible]			77.076	364	77.076	364		
TOTAL		13.[illegible]19		3.043.09[illegible]	1[illegible]3.989	77.076	364	3.602.543	158.948

Table 28 - Continued

MATERIAL BALANCE FOR BUREAU OF MINES/SELEXOL SYSTEM

STREAM		9		10		11		12	
	M.W.	LB/HR	MOL/HR	LB/HR	MOL/HR	LB/HR	MOL/HR	LB/HR	MOL/HR
N_2	28.02								
CO	28.01								
CO_2	44.01								
H_2	2.016								
CH_4	16.04								
H_2S	34.08								
COS	60.08								
NH_3	17.03								
H_2O	18.02	804593	44650	168073	9327	168073	9327	636520	35323
TAR	212			77076	364				
TOTAL		804593	44650	245149	9691	168073	9327	636520	35323

STREAM		13		14		15		16	
	M.W.	LB/HR	MOL/HR	LB/HR	MOL/HR	LB/HR	MOL/HR	LB/HR	MOL/HR
N_2	28.02			1617931	57742	1617931	57742		
CO	28.01			743637	26549	743637	26549		
CO_2	44.01			294691	6696	291742	6629	2949	67
H_2	2.016			35772	17744	35772	17744		
CH_4	16.04			55819	3480	55819	3480		
H_2S	34.08			27094	795	21675	636	5419	159
COS	60.08			721	12	721	12		
NH_3	17.03			13437	789	1618	95	11819	694
H_2O	18.02	167712	9307	9208	511	9136	507	1370277	76042
TAR	212								
TOTAL		167712	9307	2798310	114318	2778051	113394	1390464	76962

Table 28 - Continued

MATERIAL BALANCE FOR BUREAU OF MINES/SELEXOL SYSTEM

STREAM		25		26		27		28	
	M.W.	LB/HR	MOL/HR	LB/HR	MOL/HR	LB/HR	MOL/HR	LB/HR	MOL/HR
N_2	28.02								
CO	28.01								
CO_2	44.01					44	1	2905	66
H_2	2.016								
CH_4	16.04								
H_2S	34.08					68	2	5351	157
COS	60.08								
NH_3	17.03	11104	652	119	7	596	35		
H_2O	18.02			9244	513	165442	9181	288	16
TOTAL		11104	652	9363	520	166150	9219	8544	239

STREAM		29		30		31	
	M.W.	LB/HR	MOL/HR	LB/HR	MOL/HR	LB/HR	MOL/HR
O_2	32.00	34304	1072	6848	214		
N_2	28.02	113061	4035	129060	4606		
CO_2	44.01			67335	1530		
SO_2	64.06			2562	40		
NO	30.01			1861	62		
H_2O	18.02			22994	1276		
Sulfur	32.06					24183	754
TOTAL		147365	5107	230660	7728	24183	754

Table 28 - Continued

MATERIAL BALANCE FOR BUREAU OF MINES/SELEXOL SYSTEM

STREAM		17		18		19		20	
	M.W.	LB/HR	MOL/HR	LB/HR	MOL/HR	LB/HR	MOL/HR	LB/HR	MOL/HR
N_2	28.02			1614372	57615	3559	127		
CO	28.01			741173	26461	2465	88		
CO_2	44.01			244784	5562	46959	1067		
H_2	2.016			35717	17717	54	27		
CH_4	16.04			55322	3449	497	31		
H_2S	34.08			102	3	21573	633		
COS	60.08			481	8	240	4		
NH_3	17.03			562	33	1056	62		
H_2O	18.02	1370205	76038	180	10	1766	98	7190	399
TOTAL		1370205	76038	2692693	110858	78169	2137	7190	399

STREAM		21		22		23		24	
	M.W.	LB/HR	MOL/HR	LB/HR	MOL/HR	LB/HR	MOL/HR	LB/HR	MOL/HR
N_2	28.02	12441	444	1601931	57171				
CO	28.01	5714	204	735459	26257				
CO_2	44.01	1892	43	242891	5519	2905	66	44	1
H_2	2.016	274	136	34553	17581				
CH_4	16.04	433	27	54889	3422				
H_2S	34.08			102	3	5351	157	68	2
COS	60.08			481	8				
NH_3	17.03			562	33	11223	659	596	35
H_2O	18.02			108	10	9533	529	1535646	85219
TOTAL		20754	854	2671938	110004	29012	1411	1536354	85257

Table 29

BUREAU OF MINES/IRON OXIDE SYSTEM

EQUIPMENT LIST

ITEM	DESCRIPTION
Pumps	
P-101	Boiler Recycle Pump
P-120	Boiler Feed Water Pump
Reactors	
R-101	Sulfur Absorber/Regenerator
Drums	
D-101	Steam Drum
Exchangers	
E-101	Waste Heat Boiler
E-102	Economizer
E-103	Gas Cooler
Compressors	
C-101	Air Compressor
Separators	
F-101	Gasifier Off-Gas Cyclone
F-102	Regenerator Off-Gas Cyclone

Table 30

MATERIAL BALANCE FOR BUREAU OF MINES/IRON OXIDE SYSTEM

(See Figure 50)

STREAM		1		2		3		4	
	M.W.	LB/HR	MOL/HR	LB/HR	MOL/HR	LB/HR	MOL/HR	LB/HR	MOL/HR
C_2	32.00			438240	13695				
N_2	28.02			1443366	51512				
CO	28.01								
CO_2	44.01								
H_2	2.016								
CH_4	16.04								
H_2S	34.08								
COS	60.08								
NH_3	17.03								
H_2O	18.02					244045	13543		
TAR	212								
ASH								114132	
TOTAL		700000		1881606	65207	244045	13543	114132	

STREAM		5		6		7		8	
	M.W.	LB/HR	MOL/HR	LB/HR	MOL/HR	LB/HR	MOL/HR	LB/HR	MOL/HR
N_2	28.02			1443366	51512	1342382	47908	337865	12058
CO	28.01			639076	22816	431466	15404		
CO_2	44.01			253278	5755	491900	11177		
H_2	2.016			27000	13393	36830	18269		
CH_4	16.04			47960	2990	44607	2781		
H_2S	34.08			25867	759	307	9		
COS	60.08			721	12				
NH_3	17.03			11649	684	10831	636		
H_2O	18.02			175136	9719	70656	3921		
TAR	212			73819	348	68651	324		
O_2	32.00							102528	3204
TOTAL		11625		2697872	107988	2497630	100429	440393	15262

Table 30 - Continued

MATERIAL BALANCE FOR BUREAU OF MINES/IRON OXIDE SYSTEM

STREAM	M.W.	13 LB/HR	13 MOL/HR	14 LB/HR	14 MOL/HR	15 LB/HR	15 MOL/HR	16 LB/HR	16 MOL/HR
O_2	32.00	67776	2118			59936	1873		
N_2	28.02	223291	7969	101040	3606	337865	12058		
CO	28.01			32492	1160				
CO_2	44.01			37012	841				
H_2	2.016			2772	1375				
CH_4	16.04			3352	209				
H_2S	34.08			34	1				
COS	60.08								
SO_2	64.06					48750	761		
NH_3	17.03			817	48				
H_2O	18.02			5316	295			141148	7833
TAR	212			5168	24				
TOTAL		291067	10087	188003	7559	446551	14692	141148	7833

STREAM	M.W.	17 LB/HR	17 MOL/HR	18 LB/HR	18 MOL/HR	19 LB/HR	19 MOL/HR
H_2O	18.02	138376	7679	2775	154		
Na_2SO_3	126.05					7185	57
TOTAL		138376	7679	2775	154	7185	57

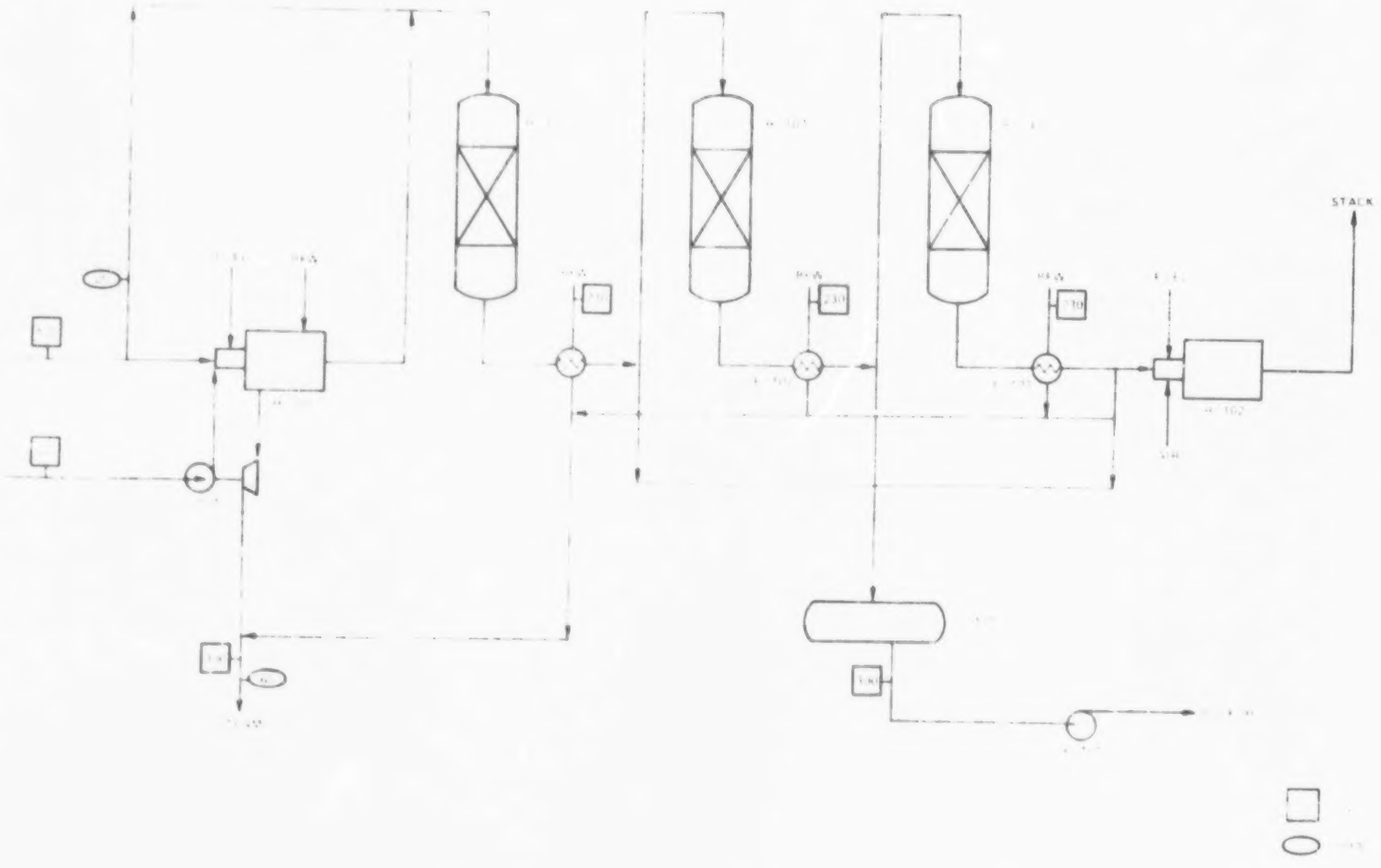
Table 30 - Continued

MATERIAL BALANCE FOR BUREAU OF MINES/IRON OXIDE SYSTEM

STREAM		9		10		11		12	
	M. W.	LB/HR	MOL/HR	LB/HR	MOL/HR	LB/HR	MOL/HR	LB/HR	MOL/HR
O_2	32.00			77632	2426				
N_2	28.02	1443422	51514	662169	23632				
CO	28.01	463958	16564						
CO_2	44.01	528912	12018	113194	2572				
H_2	2.016	39602	19644						
CH_4	16.04	47960	2990						
H_2S	34.08	341	10						
COS	60.08								
SO_2	64.06			4292	67				
NH_3	17.03	11649	684						
NO	30.01			1621	54				
H_2O	18.02	75972	4216	43969	2440				
Sulfur	32.06					20518	640		
TAR	212	73819	348						
DUST								1994	
TOTAL		2685635	107988	902877	31191	20518	640	1994	

FIG. 51

PROCESS FLOW DIAGRAM
SULFUR RECOVERY UNIT CLAUS PROCESS


STACK
BFW
230

About 3 percent of the coal heating value is removed in the gasifier cooling jacket via generation of 1315 psia saturated steam. This internally generated steam supplies 50 percent of the steam required for the gasification reaction.

Bureau of Mines/Iron Oxide System Process Description - In this system, coal is the only carbonaceous feed to the gasifier. Unreacted tar in the raw producer gas is not condensed at the cleanup system operating temperature and, therefore, cannot be recycled. Reaction of the coal with 800 F air and 1315 psia saturated steam results in about 84 percent gasification of the coal carbon with the majority of the remaining carbon released as a component of the tar. Raw producer gas leaves the gasifier at 1000 F and 400 psia and contains 1.5 lb tar per 1000 SCF and 0.3 lb dust per 1000 SCF. The gas passes first through a cyclone separator and then through an undefined high-temperature particulate removal system to reduce the dust loading to an acceptable level. Desulfurization takes place over a fixed bed of sintered iron oxide/fly ash absorbent in R-101 at a space velocity of 1000 reciprocal hours (measured at 60 F and 1.0 atm). A 90 percent approach to shift equilibrium and 100 percent approach to sulfur removal is assumed giving a residual sulfur content in the treated gas of 100 ppm. The overall desulfurization reactions are exothermic so that the fuel gas product is delivered to battery limits at 1070 F and 355 psia. When the sulfur loading reaches 6 lb/100 lb absorbent, the absorption cycle is discontinued and the spent absorbent is regenerated with air at a maximum bed temperature of 1500 F. Regeneration proceeds at a high rate, 6-10 percent SO_2 in the gas stream, for the first 30 minutes and decreases to about 1 percent SO_2 after one hour. During the one hour regeneration period 80 percent of the sulfur is recovered as SO_2 with the off-gas having an average concentration of 4 percent by volume SO_2. The regenerator off-gas is cooled to 150 F via generation of 65 psia steam in waste heat boiler E-101, boiler feed water preheater E-102, and air fan cooler E-103.

Recovery of elemental sulfur from the off-gas is complicated by the fact that sulfur is present as SO_2 in variable concentration. The SO_2 is first concentrated by absorption in an alkali solution from which it is steam stripped. About 93 percent of the feed SO_2 is recovered as a 50 percent by volume SO_2 stream. Two-thirds of the recovered SO_2 is converted to H_2S using a portion of the product fuel gas as the reducing agent. The remaining SO_2 is then reacted with the H_2S to form elemental sulfur in a typical vapor phase Claus unit. Overall recovery of elemental sulfur present in the coal is 84 percent, and 7 percent of the fuel gas product is used in the sulfur recovery section. Of the sulfur lost, 1.3 percent is in the powerplant stack, 8.7 percent in the Claus plant exhaust and 7.0 percent in the Claus purge stream. The 65 psia steam generated in E-101 represents 20 percent of the low-pressure steam requirements for the SO_2 concentration unit.

The tar present in the fuel gas product has the following analysis:

	C	H	O	N	S
% Wt.	84.02	7.77	4.66	2.00	1.55

Second-Generation Standard Systems

Gasifier/cleanup system combinations evaluated for application in second-generation integrated power systems were:

(1) BCR/Selexol
(2) BCR/Consol

The two-stage BCR gasification system operates on a 350 tons per hour of Illinois No. 6 coal having the following analysis:

	C	H	S	O	N	ASH	H_2O
Wt. %	67.4	5.1	3.8	9.6	1.2	8.7	4.2

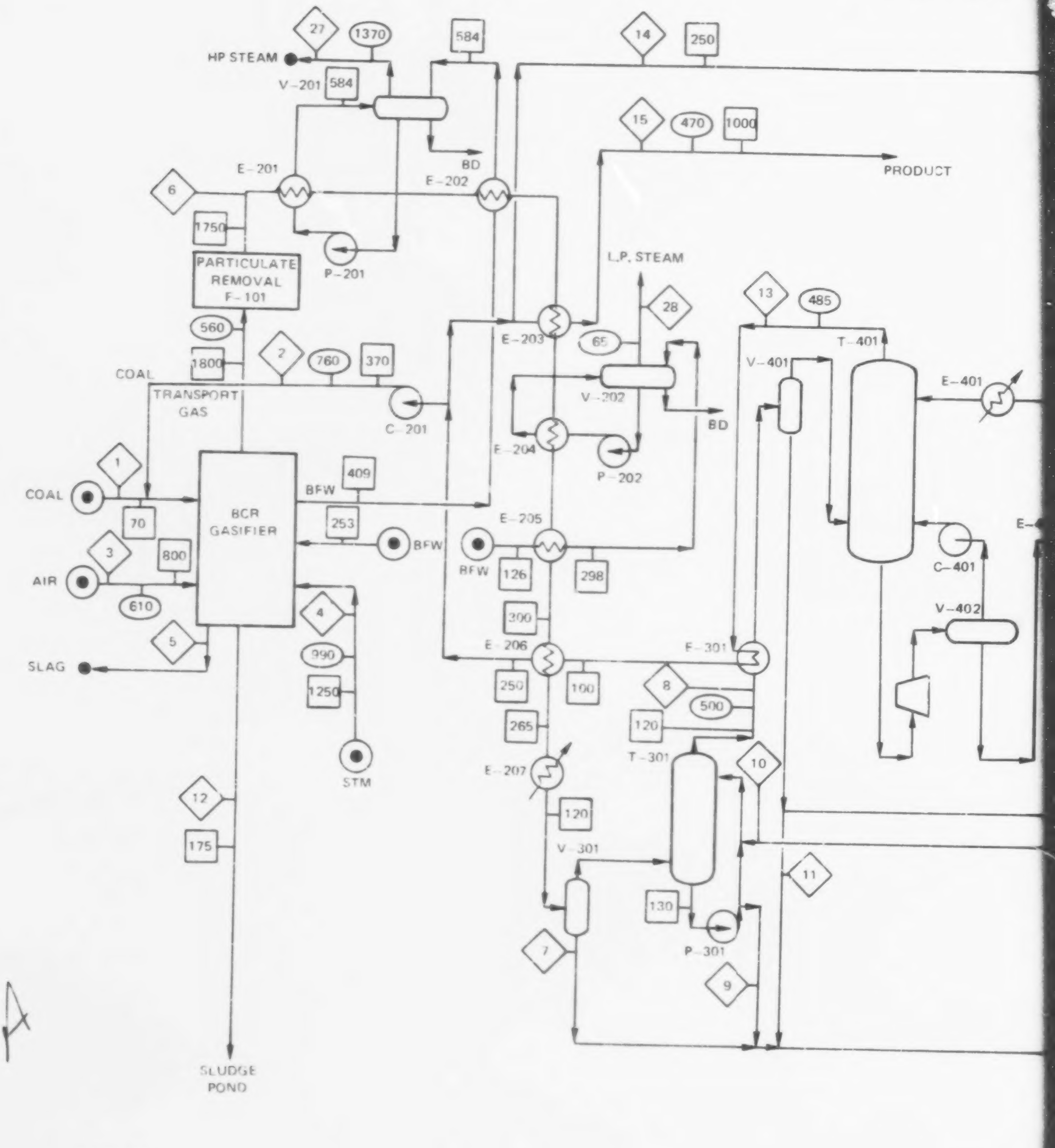
HHV = 12200 Btu/lb

Operating conditions for the gasifier were 3.1 lb air/lb coal, 0.57 lb steam/lb coal with first stage and second stage exit temperatures of 2800 F and 1800 F, respectively. A portion of the clean fuel gas product, 6.6 SCF/lb of coal, is recycled to the gasifier to transport the pulverized coal into the second stage. About 95 percent of the coal nitrogen is converted to ammonia and the COS content in the raw gas corresponds to the equilibrium for the hydrolysis reaction to H_2S. Heat is recovered from recycle char cooling in the amount of 190 Btu/lb of coal and is available for steam generation and boiler feed water heating.

Process flow diagrams for the low- and high-temperature gas purification schemes are given in Figures 52 and 53, respectively. Corresponding process equipment lists and material balance summaries are given in Tables 31 through 34.

BCR/Selexol System - Process Description - Coal is gasified using 600 F air and 1250 psia superheated steam. Under the reaction conditions, essentially complete carbon conversion is obtained. Molten slag drained from the bottom stage is water quenched and sent to battery limits for disposal. The slag quench water is clarified and sent to a sludge pond for additional solids removal before being reused. Raw producer gas leaving the top stage of the

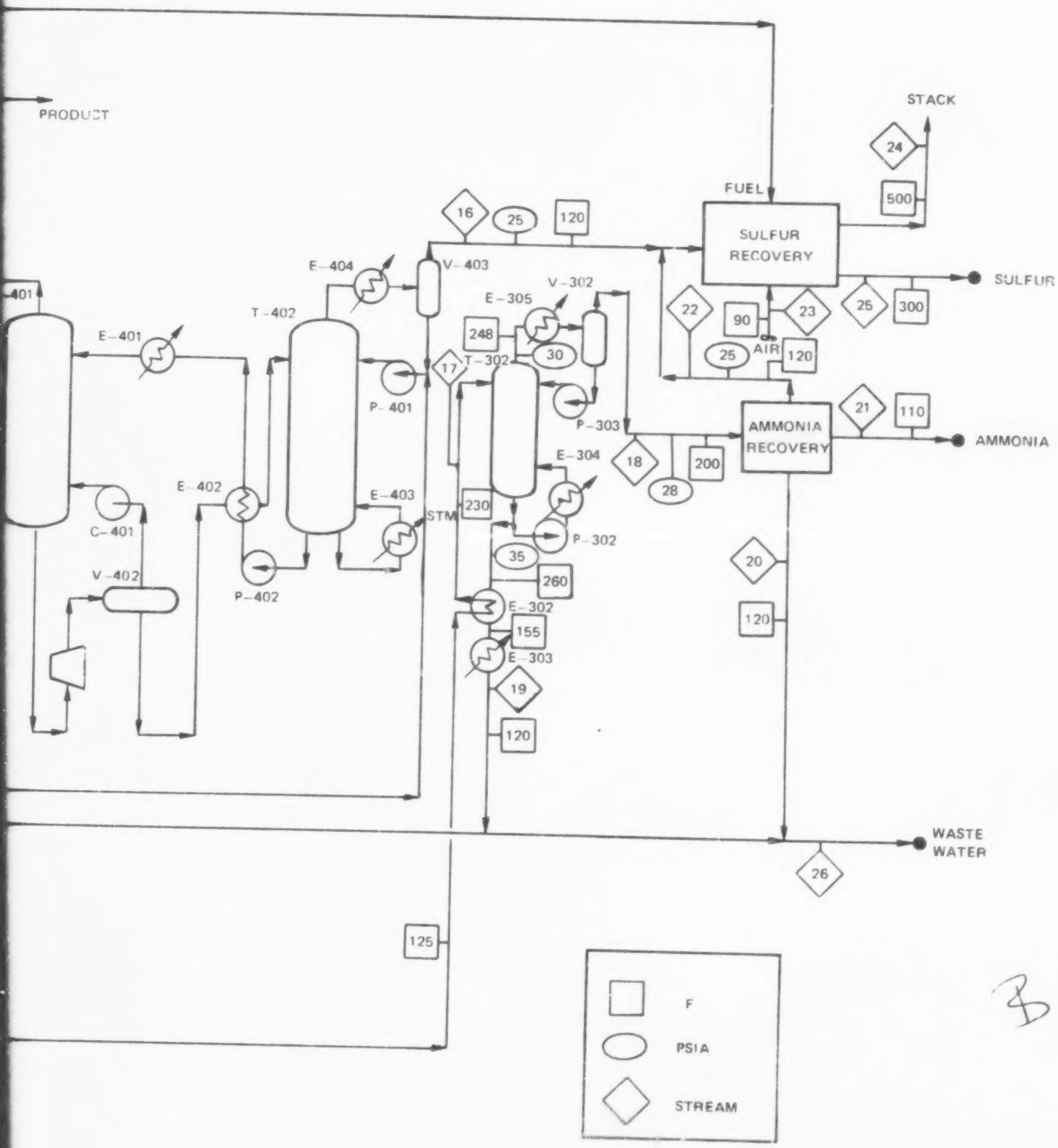
PROCESS FLOW DIAG
BCR/SELEXOL SYST


HP STEAM
27
1370
584
V-201
584
BD
14
250
15
470
1000
PRODUCT
E-201
E-202
6
1750
P-201
PARTICULATE REMOVAL F-101
L.P. STEAM
28
13
485
560
E-203
65
T-401
V-401
2
760
370
1800
COAL
TRANSPORT GAS
C-201
V-202
BD
E-401
E-204
P-202
1
COAL
70
BCR GASIFIER
BFW
409
253
BFW
E-205
BFW
126
298
C-401
3
800
AIR
610
4
300
V-402
990
E-206
E-301
5
SLAG
1250
250
100
8
500
120
265
STM
T-301
10
E-207
12
120
175
V-301
11
130
P-301
7
9
SLUDGE POND

R1-19-25

ROCESS FLOW DIAGRAM
BCR/SELEXOL SYSTEM

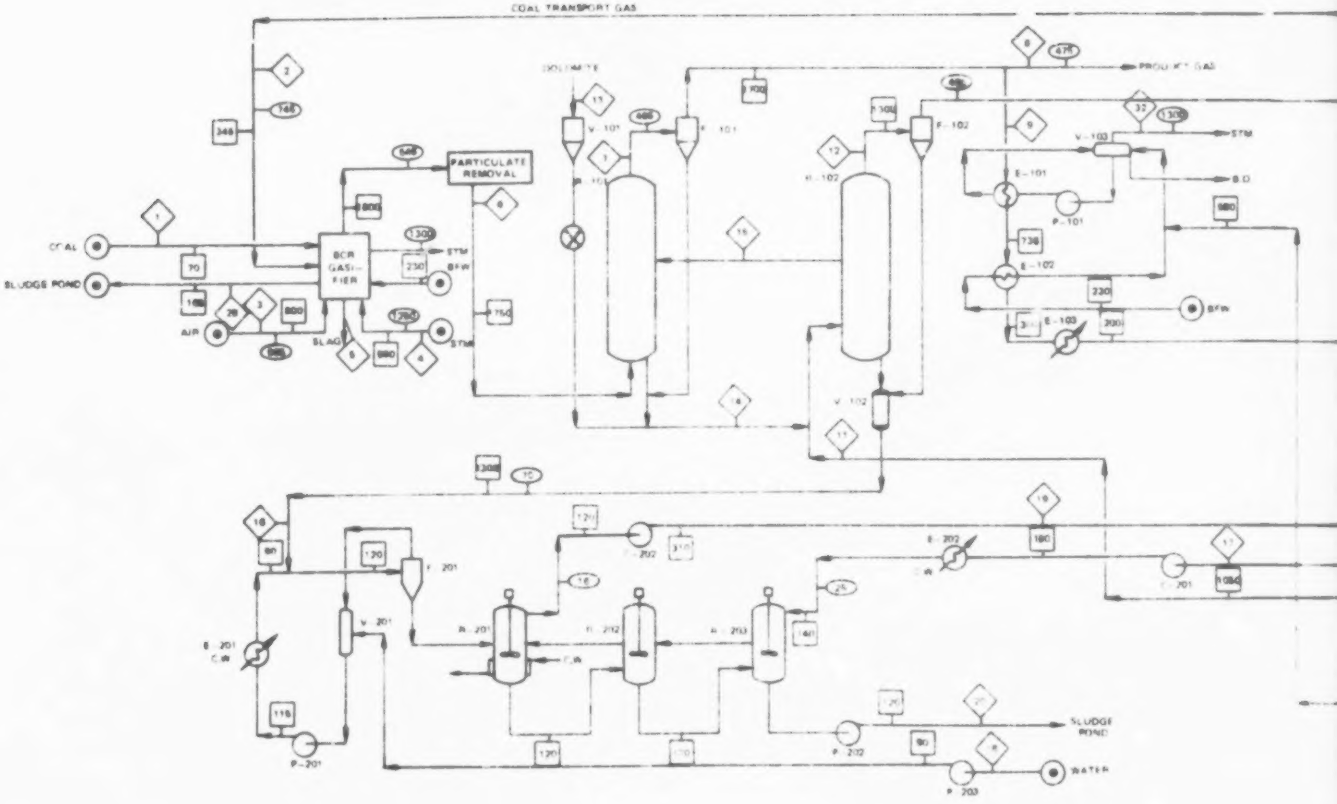
FIG. 52



PRODUCT
STACK
24
500
FUEL
SULFUR RECOVERY
SULFUR
25
300
16
25
120
V-403
E-404
401
T-402
E-401
V-302
E-305
248
30
22
90
23
AIR
120
25
17
T-302
P-401
P-303
AMMONIA RECOVERY
21
110
AMMONIA
18
28
200
E-304
E-402
E-403
STM
230
C-401
P-302
35
260
V-402
P-402
E-302
155
E-303
19
120
20
120
WASTE WATER
26
125
F
PSIA
STREAM

B

PROCESS FLOW DIAGRAM

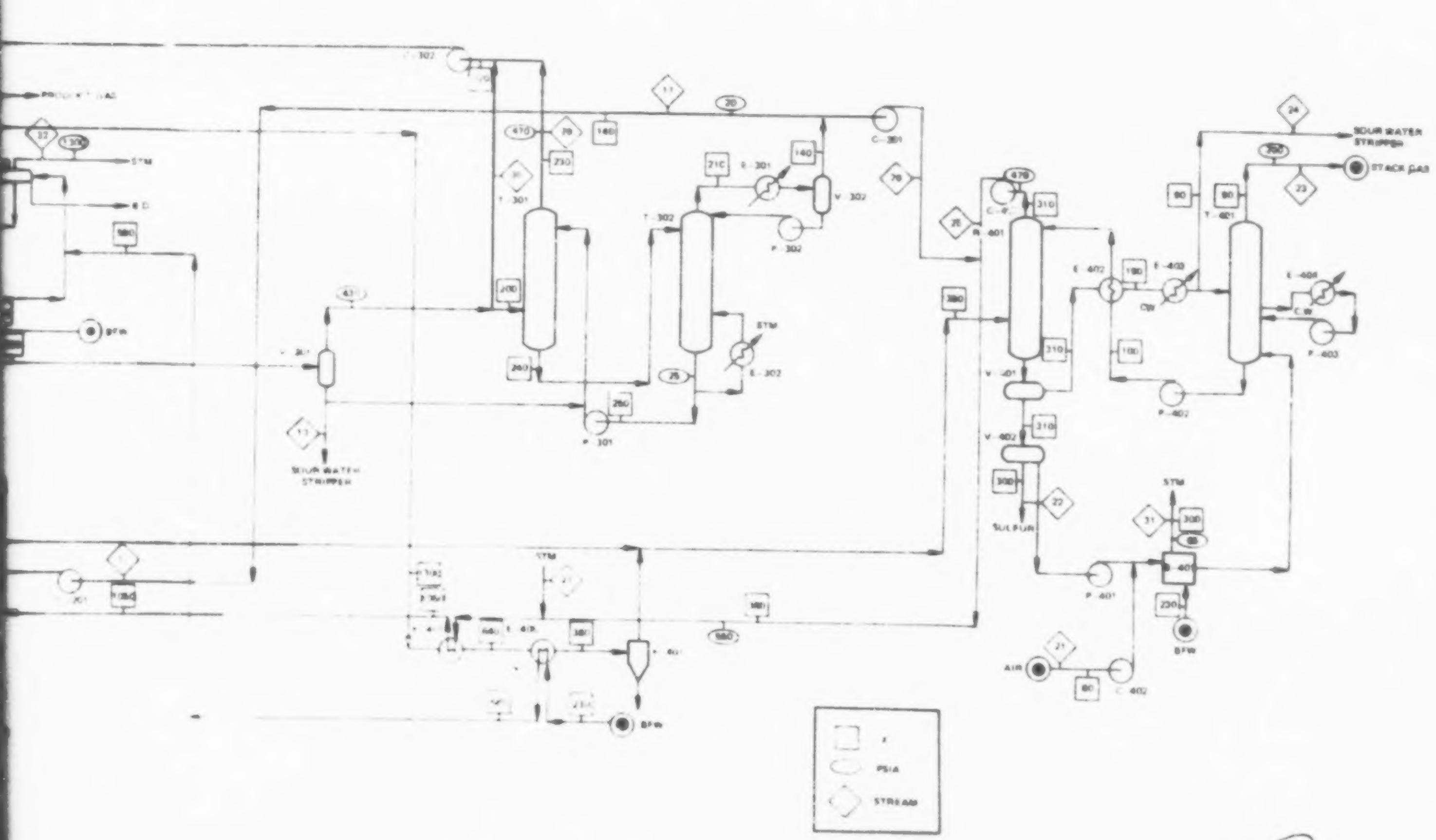


COAL TRANSPORT GAS
DOLOMITE
PRODUCT GAS
STM
B.D
BFW
COAL
SLUDGE POND
AIR
SLAG
BCR GASI-FIER
PARTICULATE REMOVAL
V-101
F-101
R-101
R-102
F-102
V-103
E-101
P-101
E-102
E-103
V-102
F-201
V-201
E-201 C.W
P-201
R-201
R-202
R-203
P-202
E-202 C.W
P-203
WATER
SLUDGE POND
C.W

A

Fig. 53

SS FLOW DIAGRAM BCR/CONSOL SYSTEM


PRODUCT GAS
32
1300
STM
B.D.
980
BFW
V-301
431
11
SOUR WATER STRIPPER
T-302
470
70
140
230
T-301
200
260
P-301
280
11
20
C-301
210
E-301
140
V-302
P-302
T-302
STM
E-302
25
70
25
R-401
380
C-401
470
310
310
V-401
310
V-402
300
22
SULFUR
E-402
180
E-403
CW
100
P-402
24
SOUR WATER STRIPPER
250
STACK GAS
23
90
90
T-401
E-404
C.W.
P-403
STM
31
300
65
E-401
P-401
230
BFW
AIR
21
80
C-402
201
1080
STM
21
660
E-401
380
980
380
BFW
F
PSIA
STREAM

B

Table 31
BCR/SELEXOL SYSTEM

EQUIPMENT LIST

SECTION 100 - GASIFICATION

ITEM	DESCRIPTION
F-101	Particulate Removal System

SECTION 200 - HEAT RECOVERY

ITEM	DESCRIPTION
Vessels	
V-201	HP Steam Drum
V-202	LP Steam Drum
Exchangers	
E-201	HP Waste-Heat Boiler
E-202	HP Economizer
E-203	Main Regenerator
E-204	LP Waste-Heat Boiler
E-205	LP Economizer
E-206	Auxillary Regenerator
E-207	Gas Cooler
Pumps	
P-201	HP Recirculating Pump
P-202	LP Recirculating Pump
Compressors	
C-201	Transport Gas Compressor

Table 31 - Continued

BCR/SELEXOL SYSTEM

EQUIPMENT LIST

SECTION 300 - GAS SCRUBBING AND SWS

ITEM	DESCRIPTION
Towers	
T-301	NH_3 Scrubber
T-302	NH_3 Stripper
Vessels	
V-301	Condensate Knock-Out Drum
V-302	NH_3 Stripper OVHD Accumulator
Exchangers	
E-301	NH_3 Scrubber OVHD Exchanger
E-302	NH_3 Stripper BTMS Exchanger
E-303	NH_3 Stripper BTMS Cooler
E-304	NH_3 Stripper Reboiler
E-305	NH_3 Stripper OVHD Condenser
Pumps	
P-301	NH_3 Absorber BTMS Pump
P-302	NH_3 Stripper BTMS Pump
P-303	NH_3 Stripper Reflux Pump

Table 31 - Continued

BCR/SELEXOL SYSTEM

EQUIPMENT LIST

SECTION 400 - ACID GAS REMOVAL

ITEM	DESCRIPTION
Towers	
T-401	Selexol Scrubber
T-402	Selexol Stripper
Vessels	
V-401	Condensate Knock-Out Drum 2
V-402	Selexol Flash Drum
V-403	Selexol Stripper OVHD Accumulator
Exchangers	
E-401	Lean Solvent Cooler
E-402	Rich/Lean Solvent Exchanger
E-403	Selexol Stripper Reboiler
E-404	Selexol Stripper OVHD Cooler
Pumps	
P-401	Selexol Stripper Reflux Pump
P-402	Selexol Stripper BTMS Pump
Compressor	
C-401	Recycle Gas Compressor

SECTION 500 - AMMONIA RECOVERY

SECTION 600 - SULFUR RECOVERY

Table 2

MATERIAL BALANCE FOR BCR/SELEXOL SYSTEM

(see Figure 52)

STREAM	M.W.	LB/HR	^{1}MOL/HR	LB/HR	^{2}MOL/HR	LB/HR	^{3}MOL/HR	LB/HR	^{4}MOL/HR
O_2	32.00					503392	15731.0		
N_2	28.02			179914	6420.9	1658190	59178.8		
CO	28.01			71809	2563.7				
CO_2	44.01			34491	783.7				
H_2	2.016			3621	1795.9				
CH_4	16.04			8152	508.2				
H_2S	34.08			3	0.1				
COS	60.08			66	1.1				
NH_3	17.03			2	0.1				
H_2O	18.02			23	1.3			396900	22025.5
TOTAL		700000		298081	12075.0	2161582	74909.8	396900	22025.5

STREAM	M.W.	LB/HR	^{5}MOL/HR	LB/HR	^{6}MOL/HR	LB/HR	^{7}MOL/HR	LB/HR	^{8}MOL/HR
N_2	28.02			1835271	65498.6			1835271	65498.6
CO	28.01			732492	26151.1			732492	26151.1
CO_2	44.01			517162	11751.0			511990	11633.5
H_2	2.016			36932	18319.4			36932	18319.4
CH_4	16.04			83222	5188.4			83222	5188.4
H_2S	34.08			23416	687.1			21075	618.4
COS	60.08			8549	142.3			8549	142.3
NH_3	17.03			9763	573.3			976	57.3
H_2O	18.02			251438	13953.3	243784	13528.5	7846	435.4
TOTAL		60000		3498245	142264.5	243784	13528.5	3238353	128044.4

Table 32 - Continued

MATERIAL BALANCE FOR BCR/SELEXOL SYSTEM

STREAM	M.W.	LB/HR	^{9}MOL/HR	LB/HR	^{10}MOL/HR	LB/HR	^{11}MOL/HR	LB/HR	^{12}MOL/HR
N_2	28.02								
CO	28.01								
CO_2	44.01	5171	117.5						
H_2	2.016								
CH_4	16.04								
H_2S	34.08	2341	68.7						
COS	60.08								
NH_3	17.03	8787	516.0						
N_2O	18.02	1017521	56466.2	1017712	56476.8	1710	94.9	210000	11653.7
TOTAL		1033820	57168.4	1017712	56476.8	1710	94.9	210000	11653.7

STREAM	M.W.	LB/HR	^{13}MOL/HR	LB/HR	^{14}MOL/HR	LB/HR	^{15}MOL/HR	LB/HR	^{16}MOL/HR
N_2	28.02	1835271	65498.6	32265	1151.5	1623092	57926.2		
CO	28.01	732492	26151.1	12876	459.7	647807	23127.7		
CO_2	44.01	351838	7994.5	6183	140.5	311164	7070.3	160152	3639.0
H_2	2.016	36932	18319.4	649	322.1	32662	16201.4		
CH_4	16.04	83155	5184.2	1461	91.1	73542	4584.9	67	4.2
H_2S	34.08	41	1.2			37	1.1	21034	617.2
COS	60.08	655	10.9	12	0.2	577	9.6	7895	131.4
NH_3	17.03	22	1.3			20	1.2	950	56.0
H_2O	18.02	234	13.0	4	0.2	207	11.5	5902	327.5
TOTAL		3040640	123174.2	53450	2165.3	2689108	108933.9	196004	4775.3

Table 32 - Continued

MATERIAL BALANCE FOR RCR/SELEXOL SYSTEM

STREAM	M.W.	17 LB/HR	17 MOL/HR	18 LB/HR	18 MOL/HR	19 LB/HR	19 MOL/HR	20 LB/HR	20 MOL/HR
N_2	28.02								
CO	28.01								
CO_2	44.01	517[illegible]	11[illegible].5	5118	11[illegible].[illegible]	5[illegible]	1.2		
H_2	2.016								
CH_4	16.04								
H_2S	34.08	2341	[illegible].7	2317	[illegible].[illegible]	24	0.7		
COS	60.08								
NH_3	17.03	[illegible]	51[illegible].	8348	4[illegible].[illegible]	4[illegible]	25.[illegible]	83	4.9
H_2O	18.02	12[illegible]2015	[illegible]	72[illegible]7	404.4	1255[illegible]	[illegible]685.2	7055	391.5
TOTAL		12[illegible]4	[illegible]	23070	1078.[illegible]	1256243	[illegible]712.9	7138	396.4

STREAM	M.W.	21 LB/HR	21 MOL/HR	22 LB/HR	22 MOL/HR	23 LB/HR	23 MOL/HR	24 LB/HR	24 MOL/HR
O_2	32.00							9002	281.3
N_2	28.02					45011	1406.[illegible]	180536	6443.1
CO_2	44.01			511[illegible]	11[illegible].[illegible]	[illegible]	5291.6	201671	4582.4
NO	30.01							1681	56.0
H_2S	34.08			231[illegible]	[illegible].[illegible]				
SO_2	64.[illegible]							41[illegible]	[illegible]5.5
NH_3	17.03	[illegible]5	4[illegible]5.						
H_2O	18.02			23[illegible]	12.[illegible]			2923[illegible]	1622.5
TOTAL		[illegible]5	4[illegible]5.[illegible]	7667	19[illegible].[illegible]	[illegible]	[illegible].2	426323	13050.8

Table 32 - Continued

MATERIAL BALANCE FOR BCR/SELEXOL SYSTEM

STREAM		25		26		27		28	
	M.W.	LB/HR	MOL/HR	LB/HR	MOL/HR	LB/HR	MOL/HR	LB/HR	MOL/HR
Sulfur	32.06	24086	751.3						
CO_2	44.01			53	1.2				
H_2S	34.08			24	0.7				
NH_3	17.03			439	25.8				
H_2O	18.02			238015	13208.4	754920	41893.4	469270	26041.6
TOTAL		24086	751.3	238531	13236.1	754920	41893.4	469270	26041.6

Table 33

BCR/CONSOL SYSTEM

EQUIPMENT LIST

SECTION 100 - DESULFURIZATION

ITEM	DESCRIPTION
Reactors	
R-101	Sulfur Absorber
R-102	Acceptor Regenerator
Vessels	
V-101	Dolomite Feed Hopper
V-102	Spent Dolomite Hopper
V-103	HP Steam Drum
Pumps	
P-101	BRW Circulation Pump
Exchangers	
E-101	Waste Heat Boiler
E-102	Economizer
E-103	Gas Cooler
Miscellaneous	
F-101	Absorber Cyclone Separator
F-102	Regenerator Cyclone Separator

Table 33 - Continued

BCR/CONSOL SYSTEM

EQUIPMENT LIST

SECTION 200 - SPENT DOLOMITE TREATING

ITEM	DESCRIPTION
Reactors	
R-201	Acceptor Converter 1st Stage
R-202	Acceptor Converter 2nd Stage
R-203	Acceptor Converter 3rd Stage
Vessels	
V-201	Quench Water Surge
Pumps	
P-201	Quench Water Pump
P-202	Dolomite Slurry Pump
P-203	Make-up Water Pump
Exchangers	
E-201	Quench Water Cooler
E-202	CO_2 Trim Cooler
Compressors	
C-201	CO_2 Blower
C-202	Acid Gas Compressor
Miscellaneous	
F-201	Hydroclone

Table 33 - Continued

BCR/CONSOL SYSTEM

EQUIPMENT LIST

SECTION 300 - CO_2 RECOVERY

ITEM	DESCRIPTION
Tower	
T-301	CO_2 Absorber
T-203	CO_2 Stripper
Vessels	
V-301	Water Separating Drum
V-302	Stripper OVHD Accumulator
Pumps	
P-301	Stripper BTMS Pump
P-302	Stripper Reflux Pump
Compressors	
C-301	CO_2 Blower
C-302	Transport Gas Compressor
Exchangers	
E-301	Stripper OVHD Condenser
E-302	Stripper Reboiler

Table 33 - Continued

BCR/CONSOL SYSTEM

EQUIPMENT LIST

SECTION 400 - SULFUR RECOVERY

ITEM	DESCRIPTION
Reactors	
R-401	Liquid Phase Clause Reactor
Towers	
T-401	SO_2 Absorption Column
Vessels	
V-401	Sulfur Separator Drum
V-402	Sulfur Storage Drum
Pumps	
P-401	Sulfur Pump
P-402	Acid Pump
P-403	Acid Circulating Pump
Compressors	
C-401	Recycle CO_2 Compressor
C-402	Air Compressor
Exchangers	
E-401	Recycle CO_2 Reheater
E-402	Feed/Bottoms Exchanger
E-403	Weak Acid Cooler
E-404	SO_2 Absorber Intercooler
E-405	BFW Preheater
Miscellaneous	
F-401	Electrostatic Precipitator
B-401	Sulfur Burner

Table 34

MATERIAL BALANCE FOR BCR/CONSOL SYSTEM

(see Figure 53)

STREAM	M.W.	LB/HR	1 MOL/HR	LB/HR	2 MOL/HR	LB/HR	3 MOL/HR	LB/HR	4 MOL/HR
O_2	32.00					503395	15731.1		
N_2	28.02			178053	6354.5	1658187	59178.7		
CO	28.01			67395	2406.1				
CO_2	44.01			23537	534.8				
H_2	2.016			3761	1865.7				
CH_4	16.04			7919	493.7				
H_2S	34.08			228	6.7				
COS	60.08			54	0.9				
NH_3	17.03			1003	58.9				
H_2O	18.02			6374	353.7			396900	22025.5
TOTAL		700000		288324	12075.0	2161582	74909.8	396900	22025.5

STREAM	M.W.	LB/HR	5 MOL/HR	LB/HR	6 MOL/HR	LB/HR	7 MOL/HR	LB/HR	8 MOL/HR
N_2	28.02			1839079	65634.5	1839079	65634.5	1661026	59280.0
CO	28.01			716557	25582.2	696085	24851.3	628693	22445.3
CO_2	44.01			513583	11669.7	584880	13289.7	528252	12003.0
H_2	2.016			37374	18538.8	38850	19270.9	35089	17405.1
$CH_4$1	16.04			81796	5099.5	81796	5099.5	73877	4605.8
H_2S	34.08			23369	685.7	2341	68.7	2113	62.0
COS	60.08			8621	143.5	517	8.6	469	7.8
NH_3	17.03			10371	609.0	10371	609.0	9367	550.0
H_2O	18.02			258373	14338.1	256288	14222.4	231474	12845.4
TOTAL		60900		3489123	142301.0	3510207	143054.5	3170360	129204.4

Table 34 - Continued

MATERIAL BALANCE FOR BCR/CONSOL SYSTEM

STREAM		9		10		11		12	
	M.W.	LB/HR	MOL/HR	LB/HR	MOL/HR	LB/HR	MOL/HR	LB/HR	MOL/HR
N_2	28.02	178053	6354.5						
CO	28.01	67392	2406.0						
CO_2	44.01	56628	1286.7			311071	7068.2	282443	6417.7
H_2	2.016	3761	1865.7						
CH_4	16.04	7919	493.7						
H_2S	34.08	228	6.7			2846	83.5	26985	791.8
COS	60.08	48	0.8						
NH_3	17.03	1005	59.0						
H_2O	18.02	24814	1377.0	16142	895.8	70588	3917.2	57824	3208.9
TOTAL		339848	13850.1	16142	895.8	384505	11068.9	367252	10418.4

STREAM		13		14		15		16	
	M.W.	LB/HR	MOL/HR	LB/HR	MOL/HR	LB/HR	MOL/HR	LB/HR	MOL/HR
$CaCO_3MgCO_3$	184.01	10636	57.8	10636	57.8				
$CaCO_3MgO$	140.41			94201	670.9	199775	1422.8	1994	14.2
CaS MgO	112.46			574997	5112.9	490438	4361.0	4903	43.6
INERTS	100	1010	10.1	101750	1017.5	100740	1007.4	1010	10.1
TOTAL		11646	67.9	781584	6859.1	790953	6791.2	7907	67.9

Table 34 - Continued

MATERIAL BALANCE FOR BCR/CONSOL SYSTEM

STREAM	M.W.	LB/HR	^{17}MOL/HR	LB/HR	^{18}MOL/HR	LB/HR	^{19}MOL/HR	LB/HR	^{20}MOL/HR
CO_2	44.01	5580	126.8			1118	25.4		
H_2S	34.08					1486	43.6		
H_2O	18.02	387	21.5	14780	820.2	148	8.2	14234	789.9
$CaCO_3MgCO_3$	184.01							10636	57.8
INERTS	100							1010	10.1
TOTAL		5967	148.3	14780	820.2	2752	77.2	25880	857.8

STREAM	M.W.	LB/HR	^{21}MOL/HR	LB/HR	^{22}MOL/HR	LB/HR	^{23}MOL/HR	LB/HR	^{24}MOL/HR
O_2	32.00	13312	416.0			1210	37.8		
N_2	28.02	43851	1565.0			43851	1565.0		
SO_2	64.06					288	4.5		
H_2O	18.02					101	5.6	48436	2687.9
Sulfur	32.06			23962	747.4				
TOTAL		57163	1981.0	23962	747.4	45450	1612.9	48436	2687.9

Table 34 - Continued

MATERIAL BALANCE FOR BCR/CONSOL SYSTEM

STREAM		25		26		27		28	
	M.W.	LB/HR	MOL/HR	LB/HR	MOL/HR	LB/HR	MOL/HR	LB/HR	MOL/HR
CO_2	44.01	283561	6443.1	27511	625.1				
H_2S	34.08	2846	83.5						
H_2O	18.02	22903	1271.0	1910	106.0	45774	2540.2	210000	11653.7
TOTAL		309310	7797.6	29421	731.1	45774	2540.2	210000	11653.7

STREAM		29		30		31		32	
	M.W.	LB/HR	MOL/HR	LB/HR	MOL/HR	LB/HR	MOL/HR	LB/HR	MOL/HR
N_2	28.02	122414	4363.8	55639	1985.7				
CO	28.01	46334	1654.2	21061	751.9				
CO_2	44.01	5840	132.7	17696	402.1				
H_2	2.016	2586	1282.7	1175	583.0				
CH_4	16.04	5444	339.4	2475	154.3				
H_2S	34.08	157	4.6	72	2.1				
COS	60.08	36	0.6	18	0.3				
NH_2	17.03	690	40.5	313	18.4				
H_2O	18.02	4646	257.8	1728	95.9	48880	2712.5	193530	10739.7
TOTAL		188147	8081.3	100177	3993.7	48880	2712.5	193530	10739.7

gasifier at 1800 F and 560 psia passes through a series of cyclone separators to remove entrained char and is then cooled to 120 F in the regenerative exchangers E-203 and E-206, generating 1370 psia and 65 psia steam in the waste heat boilers E-201 and E-204, and preheating boiler feed water in E-202 and E-205. Cooled gas is scrubbed with water in T-301 to remove 90 percent of the ammonia and is then sent to the Selexol acid-gas removal system. Condensate from separator drums V-301 and V-401 is combined with the bottoms from the water scrubber and fed to the sour water stripper, T-302, where steam stripping removes 95 percent of the dissolved ammonia and essentially all of the acid gases. Part of the stripped water is recycled to the water scrubber, T-301, and the remainder is delivered to battery limits as waste water.

The overhead gas from the sour water stripper is fed to an ammonia recovery unit where 99 percent of the ammonia is recovered in the anhydrous form via absorption in an aromatic hydrocarbon solvent.

Ammonia-free producer gas is desulfurized in T-401 by countercurrent contacting with Selexol solvent at subambient temperature. The treated gas, containing 100 ppm total sulfur, is reheated against raw producer gas and is delivered to battery limits at 1000 F and 475 psia. Rich Selexol solvent, from the bottom of T-401, is flashed, heated against lean solvent in E-402, and finally stripped in the regenerator T-402. The stripped acid gases from the regenerator are cooled to 120 F and combined with the off-gas from the ammonia recovery unit to provide the feed to the sulfur recovery plant. The combined acid gas stream contains 14 percent H_2S so that a split flow, three-stage vapor phase Claus system (Figure 51, previously shown) is used to recover 95 percent of the sulfur in the elemental form. About 2 percent of the product fuel gas is required as process fuel for the sulfur recovery unit.

Low-pressure steam generated in E-204 plus that exported from the Claus sulfur plant provides 98 percent of the 50 psig steam requirement for the Selexol regenerator and the sour water stripping system. The high-pressure steam generated in E-201 is delivered to the steam generator portion of the combined-cycle power system for superheating.

BCR/Consol System - Process Description - As in the low-temperature cleanup case, coal is gasified in the two-stage BCR gasifier using 800 F air and 1250 psia superheated steam. Again, essentially complete carbon conversion is achieved. Heat from the char recycle system is used to generate 1300 psia saturated steam. After high-temperature particulate removal, the raw producer gas enters the desulfurization system at 1750 F. In the fluidized-bed desulfurizer, R-101, the gas contacts half-calcined dolomite acceptor at 495 psia. The acceptor enters with 75 percent of the calcium as CaS and leaves with 88 percent in the sulfided form. Approach to equilibrium for the CO shift and sulfur absorption reactions is assumed to be 100 percent. The overall reactions occuring during desulfurization are slightly endothermic so that the treated gas exits at 1700 F and contains 540 ppm total sulfur.

Sulfided acceptor plus an amount of fresh dolomite equal to slightly more than 1 percent of the circulating solids, are transported to the fluidized-bed regenerator, R-102, by the regeneration gas having a molar ratio of carbon dioxide to steam approximately equal to 2. Regeneration is carried out at 1300 F with an 85 percent approach to H_2S equilibrium. About 14 percent of the CaS is converted and the regenerated solids are recycled by gravity to the desulfurizer. The off-gas from the regenerator contains 7% H_2S by volume and, after cooling to 380 F, is fed to a liquid-phase sulfur recovery unit.

Spent dolomite, withdrawn from the regenerator via lock hopper V-102, is treated before discharge to the environment. This stream, containing 75 percent of the calcium as CaS, is slurried in water in a hydrocyclone F-201 and then processed in a three-stage countercurrent reactor system where CO_2 is used to convert all the calcium to the carbonate form thereby rendering the stream suitable for discharge to a sludge pond. The H_2S generated in the spent dolomite system exits from R-201, is compressed, and fed to the sulfur recovery unit along with the regenerator off-gas.

The liquid phase Claus reactor, R-401, operates at 310 F and converts 90 percent of the H_2S feed to elemental sulfur. Feed gas is contacted countercurrently by a solution of H_2SO_3 whereby sulfur is formed via the reaction

$$2H_2S + H_2SO_3 \rightarrow 3S + 3H_2O \qquad (42)$$

Liquid sulfur is separated from the aqueous phase in V-401 and stored in the sulfur accumulator V-402. A third of the sulfur product is burned with air in B-401 to produce SO_2 which in turn is absorbed by contact with water in T-401. The rich H_2SO_3 solution from T-401 is heated against the aqueous phase from V-401 and fed to the sulfur converted R-401. Excess water formed by the reaction in R-401 is removed from the system and sent to a sour water stripper.

The overhead gas from the sulfur reactor is compressed, combined with makeup CO_2 and steam, and recycled to the dolomite regenerator. Makeup CO_2, required for spent dolomite treating and acceptor regeneration, is extracted from a slip stream of product fuel gas. This stream, representing 10 percent of the product gas, is cooled from 1700 F to 200 F by 1300 psia steam generation in E-101, boiler feed water heating in E-102, and air cooling in E-103. Part of the cooled gas is then processed in a hot potassium carbonate absorption/stripping system where 85 percent of the CO_2 is removed. The treated gas from the CO_2 absorber T-301, is combined with the remainder of the product gas slip stream and recycled to the BCR gasification system as coal transport gas.

Carbon dioxide taken overhead from the "hot pot" regenerator, T-302, is compressed and used as makeup CO_2.

Product fuel gas is delivered to battery limits at 1700 F and 475 psia. Low-pressure steam generated in the sulfur burner, B-401, supplies 70 percent of the 50 psig steam required for the CO_2 removal regenerator and the sour water stripping system. The 1300 psia steam generated in the coal gasification unit and gas purification system is sent to the steam generator portion of the combined-cycle power system.

Utilities and Energy Balances

The utilities summaries for the four standard gasification/cleanup system combinations are given in Tables 35 through 38. For each system, the summary gives a section-by-section breakdown of the utilities requirement.

The energy balances around the four gasification/cleanup system combinations are shown in Table 39. The basis for these balances were the higher heating value at 77 F plus the sensible heat referred to 77 F with water in the liquid state. Electric power input was calculated on the basis of equivalent coal heating value assuming 35 percent overall efficiency on the HHV. Generally, the power consumption does not appear as process stream energy since pratically all of it is used in coal pretreatment and to provide mechanical refrigeration.

Based on the total energy input to each system, the BCR/Consol system has the highest recovery of useful energy; 90.6 percent contained in the product fuel gas and export steam. The other three systems have useful energy recoveries in the range of 76-79 percent, as illustrated in Table 39. The bulk of this differential is directly related to the energy lost in process cooling.

Investment Cost

The battery limits investment costs for the integrated gasification and gas purification systems were estimated using currently available cost data. Due to the lack of actual cost data on the high-temperature desulfurization processes, investment costs were developed from conceptual plant designs for the Consol and Sintered Iron Oxide processes. Other sections were estimated from cost curves developed from published data.

The 1974 estimated investment costs for the four standard systems are summarized in Table 40. Considering the variety of sources from which the costs were developed, an accuracy of not better than ± 25 percent should be

Table 35

SUMMARY OF BOM GASIFICATION/SELEXOL DESULFURIZATION UTILITIES CONSUMPTION

	Coal Gasification	Gas Cooling	Sour Water Stripping	Ammonia Recovery,	Acid Gas Removal	Sulfur Recovery	Total
STEAM, LB/HR							
@ 65 psia				6420	106200	(68365)	44255
@ 1315 psia	119790			50710			170500
COOLING WATER GPM				7330	37090		44420
POWER, KW	10500	1286	716	1244	13495	4	27245
BFW, LB/HR	165355					75060	240415
STM COND., LB/HR				(57130)	(106200)		(163330)
CHEMICALS $/DAY				40	70	26	136

Table 36

SUMMARY OF BUREAU OF MINES/SINTERED IRON OXIDE

UTILITIES CONSUMPTION

	GASIFIER	SULFUR RECOVERY	DESULFURIZATION	TOTAL
Steam, Lb/Hr.				
65 psia		646,590	(138,400)	508,190
1315 psia	27,775			27,775
Cooling Water GPM		42		42
Power, KW	10,500	17,650	8,340	36,490
BFW, Lb/Hr.	218,435		141,100	359,535
STM Cond. Lb/Hr.		(646,590)		(646,590)
Chemicals $/Day		6,186		6,186

Table 37

SUMMARY OF BCR GASIFICATION/SELEXOL DESULFURIZATION UTILITIES CONSUMPTION

	Coal Gasification	Heat Recovery	Gas Scrubbing	Acid Gas Removal	Trans. Gas Compression	Sour Water Stripper	Sulfur Recovery	Ammonia Recovery	Total
STEAM, lb/hr									
@ 65 psia		(469270)		303500		271490	(96780)	4780	13720
@ 1370		(754920)						37750	(717170)
@ 1250 SPHT	396900								396900
COOLING WATER, gpm	10000			62950				5460	78410
POWER, kw	21000		783	33650	3190	569	5	926	60123
BFW, lb/hr	762470	473960					104730		1341160
STM. COND., lb/hr				(303500)		(271490)		(42530)	(617520)
PROCESS WATER lb/hr	210000								210000
CHEMICALS, $/day				100			26	30	156

Table 38

SUMMARY OF BCR GASIFICATION/CONSOL DESULFURIZATION
UTILITIES CONSUMPTION

	Coal Gasification	Gas Purification	Spent Dolomite Treating	CO_2 Removal	Sulfur Recovery	Sour Water Stripping	Total
STEAM, LB/HR							
@ 65 PSIA				58200	(48880)	13680	23000
@ 1300	(135530)	(147819)					(283349)
@ 1250 SPHT	396900						396900
COOLING WATER, GPM	10000		1120	390	6200		17710
POWER, KW	21000	277	155	4863	3931	29	30255
BFW, LB/HR	137190	195530			49370		382090
STM. COND., LB/HR				(58200)		(13680)	(71880)
PROCESS WATER LB/HR	210000						210000
CHEMICALS, $/DAY		1400					1400

Table 39

GASIFICATION/CLEAN-UP SYSTEMS ENERGY BALANCES

SYSTEM	BOM/SELEXOL	BOM/IRON OXIDE	BCR/SELEXOL	BCR/CONSOL
ENERGY IN. MMBTU/HR.				
Coal	8015.0	8015.0	8610.0	8610.0
Gasifier Air	378.9	338.0	388.3	388.3
Sulfur Recovery Air	0.1	0.5	0.6	0.1
Boiler Feed Water	36.8	55.0	173.4	58.5
Make-Up Water			4.8	5.2
Electric Power (Coal HHV)	265.7	355.8	586.3	295.0
Process Steam, 65 Psia	50.3	577.0	15.6	26.1
1300 Psia	193.2	31.5	574.5	574.5
TOTAL	8940.0	9372.8	10353.5	9957.7
ENERGY OUT, MMBTU/HR.				
Product Fuel Gas	6821.2	7265.0	7327.1	8705.8
Export Steam			812.6	321.0
Sulfur	97.6	82.8	97.2	96.7
Ammonia	98.5		73.3	
Stack Gas	49.8	53.5	76.2	0.3
BFW Blowdown	1.0	1.4	5.1	1.8
Process Cooling	1203.4	1054.9	1117.7	328.8
Process Heat Loss	174.5	218.2	98.1	173.5
Steam Condensate	36.4	144.2	137.7	16.0
Waste Water	7.6	0.1	32.0	22.4
Ash & Dust	226.2	226.3		
Power Consumption	93.0	124.5	205.2	103.3
Power Generation Loss	172.7	231.3	381.1	191.8
TOTAL	8981.9	9402.2	10363.3	9961.4
BALANCE %	100.46	100.31	100.09	100.03

Table 39 (Cont'd)

GASIFICATION/CLEAN-UP SYSTEM ENERGY BALANCES

SYSTEM	BCR/SELEXOL	BCR/CONSOL	BOM/SELEXOL	BOM/IRON OXIDE
Total Energy In. MMBTU/HR.	10353.5	9957.7	8940.0	9372.8
Coal In. MMBTU/HR.	8610.0	8610.0	8015.0	8015.0
% Recovery:				
Product Fuel Gas	70.77	87.43	76.30	77.51
Export Steam	7.85	3.22		
Sulfur	0.94	0.97	1.09	0.88
Ammonia	0.71		1.10	
Stack Gas	0.74		0.56	0.57
BFW Blowdown	0.05	0.02	0.01	0.02
Process Cooling	10.80	3.30	13.46	11.26
Process Heat Loss	0.95	1.74	1.95	2.33
Power Generation Loss	3.68	1.93	1.93	2.47
Steam Condensate	1.33	0.16	0.41	1.54
Waste Water	0.31	0.22	0.08	
Ash & Dust			2.53	2.41
Power Consumption	1.98	1.04	1.04	1.33
TOTAL	100.10	100.03	100.46	100.32

Table 40

SUMMARY OF INTEGRATED GASIFICATION/CLEANUP SYSTEM
INVESTMENT COSTS
(Millions of Dollars)
BASIS: 1974 Gulf Coast

System: GASIFIER CLEANUP	BOM SELEXOL	BOM IRON OXIDE	BCR SELEXOL	BCR CONSOL
Gasification	42	42	56	56
Gas Cooling	10	--	16	--
Desulfurization	16	14	20	14
Sour Water Stripping	4	--	4	1
Ammonia Recovery	7	--	6	--
Sulfur Recovery	2	7	2	6
TOTAL, MM$	81	63	104	77

expected. Of the total investment costs, the coal gasification system represents 50-70 percent. The available data on fixed-bed gasification systems costs is very sparse since no commercial plants have yet been constructed in the U.S. Reliable data for second-generation gasifiers is essentially nonexistent since these systems are only in developmental stages where only preliminary cost estimates have been made on conceptual pilot plant designs.

The only significant conclusions that can be drawn from the cost figures is that, for a given gasification system, the low-temperature cleanup scheme is more expensive, possibly by 30 percent. The costs for the desulfurization sections appear to be comparable for both the low- and high-temperature schemes, but the former scheme must also bear the additional costs associated with gas cooling, sour water stripping, and ammonia recovery.

Comparison of Standard Integrated Systems

Comparison of first- and second-generation gasification/cleanup system characteristics is given in Table 41. In terms of the percentage recovery of input energy in the product fuel gas and export steam, the second-generation BCR gasification system appears to be about 2.5 percent more efficient than the first-generation Bureau of Mines stirred-bed gasifier. For the first-generation gasification system, high-temperature gas cleanup does not offer significant advantage over low-temperature desulfurization. The performance of the sintered iron oxide process could be improved somewhat by a more efficient sulfur recovery system. The present scheme involves the use of large quantities of steam to recover and concentrate the SO_2 regenerator off-gas. In addition, 7 percent of the fuel gas product is used to reduce the SO_2 to elemental sulfur.

With second-generation gasification systems, high-temperature cleanup shows a decided advantage, 90.6 percent energy recovery versus 78.6 percent for the BCR/Selexol system. At least part of this differential is due to the more stringent desulfurization duty designed into the Selexol system. If the low-temperature systems were designed for a comparable 500 ppm total sulfur in the treated fuel gas, instead of 100 ppm, the energy requirement for stripping steam and refrigeration would be substantially lower. It should be noted that the Consol high-temperature process does not have the capability to reduce the sulfur content to less than 500 ppm under the operating conditions imposed. In this respect, low-temperature systems are more versatile.

Table 41

COMPARISON OF GASIFIER/CLEANUP SYSTEM PERFORMANCE

GASIFIER SYSTEM	BUREAU OF MINES		BCR	
CLEANUP SYSTEM	SELEXOL	IRON OXIDE	SELEXOL	CONSOL
Coal:	W. Kentucky No. 9		Illinois No. 6	
STPD	8400		8400	
MMBTU/HR.	8015		8610	
Product Fuel Gas:				
MMSCFD	1003.2	915.9	993.5	1178.3
°F	265	1070	1000	1700
PSIA	355	355	470	475
HHV TRU/SCF	160.2	167.9	159.2	137.1
LHR BTU/SCF	149.1	154.2	147.5	126.4
VOL. % N_2	51.97	47.70	53.18	45.88
CO	23.87	15.34	21.23	17.37
CO_2	5.02	11.13	6.49	9.29
H_2	15.98	18.19	14.87	13.47
CH_4	3.11	2.77	4.21	3.56
H_2S+COS	0.01	0.01	0.01	0.05
NH_3	0.03	0.63	--	0.43
H_2O	0.01	3.91	0.01	9.95
TAR	--	0.32	--	--
	100.00	100.00	100.00	100.00
Ammonia STPD	133.2	--	99.2	--
Sulfur LTPD	259.1	219.8	258.1	256.7
Slag/Ash STPD	1533.0	1533.0	730.8	730.8
Waste Water, GPM	351	--	897	129
SO_2 LB./MMBTU Coal	0.408	0.872	0.567	0.553
NO_2 LB/MMBTU Coal	0.545	4.524	0.306	2.939
Energy Recovery %				
Product Fuel Gas	76.30	77.51	70.77	87.43
Export Steam	0.00	0.00	7.85	3.22
	76.30	77.51	78.62	90.65

SECTION 4

INTEGRATION OF GASIFICATION, CLEANUP AND POWER SYSTEMS

Section 4 contains a description of the integration of those gasification and cleanup processes previously identified as most attractive into both conventional steam systems (2400 psi/1000 F/1000 F) and into an advanced power generation system having a corresponding time of maturity. First-generation systems are based on the use of the fixed-bed gasifier concept typified by the Bureau of Mines stirred-bed gasifier. Compatible high-and low-temperature cleanup systems were considered and the Selexol and sintered iron oxide processes were selected as the most suitable representatives in their respective categories. These two systems were combined with a first-generation COGAS power system having a turbine inlet temperature of 2200 F and using conventional cooling techniques.

Second-generation systems were focused on the use of a BCR-type entrained-flow gasifier. Low-temperature cleanup systems were judged to be sufficiently mature to preclude any significant developments in that technology; therefore, the Selexol process was again chosen to represent low-temperature cleanup. For the high-temperature process, a choice was made between the Consol half-calcined dolomite and the APC fully-calcined dolomite process. While there is only a small difference in estimated overall system performance. the fully calcined system tends to absorb CO_2 at temperatures below 1750 F causing a rapid increase in the fuel requirement for subsequent calcination. Therefore, the Consol process was selected as the representative high-temperature cleanup system. Both the BCR/Selexol and BCR/Consol systems were combined with a COGAS power cycle with a 2600 F turbine inlet temperature, ceramic vanes and conventionally cooled blades.

The salient characteristics of the major components of each system as they affect the overall system can be summarized as follows:

First Generation

Gasifier: Fixed - bed type - Bureau of Mines
Low temperature (1000 F outlet)
Ammonia and tar in product gas
High cold gas efficiency

Low-Temperature Cleanup - Selexol
Includes ammonia removal
Necessitates tar removal by water quench of gas
Effective H_2S removal

High-Temperature Cleanup - Sintered iron oxide
Effective H_2S removal below 1100 F
Regeneration produces SO_2
Difficult to produce elemental sulfur
No provision for ammonia removal
No water scrub for particulates

Power System - Combined Cycle
2200 F turbine inlet temperature
Conventional engine cooling
Two-pressure steam bottoming cycle

Power System - Steam
2400 psi throttle pressure
1000 F throttle temperature with reheat to 1000 F

Second Generation

Gasifier: Entrained Flow - BCR type
High temperature - two stage (1800 F outlet)
Ammonia and COS in product gas
Needs coal transport gas stream
High percentage of sensible heat in gas

Low-Temperature Cleanup - Selexol
- Includes ammonia removal
- Requires fuel gas regeneration to recover fuel gas sensible heat
- High solvent flow for COS removal
- Effective sulfur removal

High-Temperature Cleanup - Consol
- Potentially good COS removal
- Limited sulfur removal ability
- Low utility requirements
- No provision for ammonia removal
- No water scrub for particulates

Power System - Combined cycle
- 2600 F turbine inlet temperature
- Ceramic stator vanes
- Conventional turbine blade cooling
- Two-pressure steam bottoming cycle

Power System - Steam
- 2400 psi throttle pressure
- 1000 F throttle temperature with reheat to 1000 F

STEAM SYSTEMS

The performance of four different gasifier/cleanup steam power plant systems was evaluated for the configuration using a fuel gas let-down turbine as previously shown in Figure 15(b). The results of these evaluations are summarized in Table 42. Performance of the BCR gasifier-based systems can be compared directly to the coal-fired power plant characteristics given previously in Table 7 since they both are based on firing an Illinois Seam #6 coal. The Bureau of Mines gasifier uses Western Kentucky #9 coal as the feed. A boiler using that coal would have an efficiency of 86.9 percent as opposed to the 88.5 percent value used in the reference plant. The systems using the Bureau of Mines gasifier therefore should be compared to a direct-fired plant having an efficiency of 34.5 percent with stack gas cleanup while the BCR-based systems should be compared to 35.1 percent.

It is apparent from Table 42 that with the exception of the BCR-Consol system, the performance of these systems is quite unattractive when compared to the reference coal-fired plant with stack gas cleanup. The reasons for this

Table 42

Gasified Coal - Steam Cycle Performance Summary

	Gasifier-Cleanup Combination			
	BuMines/ Selexol	BuMines/ Iron Oxide	BCR/ Selexol	BCR/ Consol
Fuel System				
Coal Feed Rate - lb/hr	700,000	700,000	700,000	700,000
Clean Fuel Gas Temp. - F	265	1,070	1,000	1,700
Fuel Gas HHV - Btu/scf	160.3	165.9	159.3	135.8
Boiler & Steam Cycle				
Boiler Efficiency	.848	.852	.855	.867
Net Steam Cycle Heat Rate - Btu/kwh	7836	7836	7836	7836
Net Steam Cycle Output - MW	646.1	620.0	736.9	753.4
Let-Down Turbine				
Net Output - MW	0	43.3	27.7	166.6
Auxiliary Power - MW	50.0	58.0	87.1	58.3
Plant Performance				
Net Output Power - MW	596.1	605.3	677.5	861.7
Efficiency (HHV)	.254	.258	.269	.342

become quite clear when each of the major parts of the system are examined in detail considering both thermal and auxiliary power requirements. This was done in the preparation of the performance estimates and is presented in Table 43. In essence, this table summarizes the performance calculations and an explanation of the table requires a description of the general approach as well as the procedures used in the calculation.

Performance Calculations

To mate the gasifier/cleanup systems with a steam cycle is outwardly a simple matter requiring only a source of compressed air for the gasifier and utilization of the product gas in a steam boiler. These two factors can be considered separately.

The product gas utilization involves not only raising steam for power generation, but also for use in the cleanup process and the gasifier. To keep track of the various uses of this system heat requires an accounting system. The approach that has been chosen is to relate all process heat requirements to an equivalent fuel consumption in Btu/hour by dividing by boiler efficiency. While simple, this approach is also quite accurate, since by virtue of air preheaters, a steam system is able to use all the available fuel energy in the steam power cycle and the heat used to produce low-pressure steam for process use subtracts directly from that available to the steam cycle.

Boiler efficiency is calculated using an accepted industry method. (33) These efficiencies are generally two to three points lower than for direct coal firing due to the increased hydrogen content of the fuel gas caused by steam injection. Higher fuel gas temperature (sensible heat) increases the total heat content of the gas while the composition and, therefore, moisture losses remain constant. This improves boiler efficiency, partially offsetting the loss due to hydrogen content. The sensible heat content of the fuel is referenced to 59 F. It should be noted that even with 10 percent excess air for gas as opposed to 20 percent for coal the boiler efficiency for the fuel gas is lower than for coal. In proceeding through the system, the heat requirements of each system component (or the heat exported) are charged against (or credited to) that component as the equivalent fuel energy consumed at the appropriate boiler efficiency. The net fuel available is then used to calculate heat available to the steam cycle and net steam cycle output, using the steam cycle performance presented previously for the reference coal fired steam plant (Net Heat Rate = 7,836).

Table 43

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GASIFIED COAL STEAM CYCLE-COMPONENT PERFORMANCE CHARACTERISTICS

ENERGY (MM Btu/hr., unless otherwise noted)	[illegible]	[illegible]	[illegible]	[illegible]
Coal Energy Input (HHV)	[illegible]	[illegible]	[illegible]	[illegible]
Boiler				
Clean Fuel Gas Flow, lb/hr	[illegible]	[illegible]	[illegible]	[illegible]
Gas Temperature, F	[illegible]	[illegible]	[illegible]	[illegible]
Chemical Energy in Gas (HHV)	[illegible]	[illegible]	[illegible]	[illegible]
Sensible Heat in Gas (Above 80 F)	[illegible]	[illegible]	[illegible]	[illegible]
Total Energy in Gas (HHV)	[illegible]	[illegible]	[illegible]	[illegible]
Boiler Efficiency	[illegible]	[illegible]	[illegible]	[illegible]
Gasifier				
Clean Fuel Gas Energy Out (at boiler supply temperature)	[illegible]	[illegible]	[illegible]	[illegible]
Heat Available From Gasifier and Gas Streams				
Useful	[illegible]	[illegible]	[illegible]*	349
Rejected	(180) [illegible]*		[illegible]*	[illegible]
Process Heat Requirements				
Added to Fuel Gas Stream	[illegible]	[illegible]	[illegible]	
Input to Steam	[illegible]	[illegible]		
Gasifier Feed Preheat	[illegible]	[illegible]	[illegible]	510
Net Process Heat Used (Required less useful heat available)	[illegible]	[illegible]	[illegible]	[illegible]
Net Fuel Energy for Process Heat (heat/boiler efficiency)	[illegible]	[illegible]	[illegible]	[illegible]
Net Fuel Energy Out (Fuel out less fuel energy for process heat)	[illegible]	[illegible]	[illegible]	[illegible]
Effective Gasifier Efficiency	[illegible]	[illegible]	[illegible]	.891
Cleanup System				
Net Process Heat Required by Sulfur Recovery	[illegible]	[illegible]	[illegible]	[illegible]
Net Fuel Energy for Sulfur Process Heat	[illegible]	[illegible]	[illegible]	[illegible]
Fuel Energy Used Directly by Claus Plant		[illegible]	[illegible]	---
Total Fuel Energy for Sulfur Removal	110	1,111	495	[illegible]
Net Process Heat Required for NH_3 Removal	[illegible]	---	381	---
Net Fuel Energy for NH_3 Process Heat	[illegible]		[illegible]	
Total Fuel Energy Required for Cleanup	[illegible]	1,111	794	[illegible]
Net Fuel Energy Out of Cleanup	6,573	5,943	[illegible]	7,550
Steam Cycle				
Heat Available from Boiler	5,250	5,090	[illegible]	[illegible]
Overall Coal to Steam Efficiency	[illegible]	.634	[illegible]	.760
Net Steam Cycle Output, kW	657,[illegible]	648,290	760,082	835,375
Auxiliaries				
Gasifier Auxiliary Power, kw	(97) 10,[illegible]**	(97) 15,500	(278) 24,418	(250) 24,278
Auxiliary Power for Gas Cooling, kw	(11) 1,086			
Auxiliary Power for NH_3 Removal, kw	(20) 2,127		(25) 2,402	
Sulfur Auxiliaries, kw	(133) 14,365	(239) 21,000	(322) 35,090	(55) [illegible],381
Subtotal Auxiliary Power Gasifier & Cleanup, kw	28,[illegible]	36,500	61,910	30,659
Steam Plant Auxiliary Power, kw (3.31% of net steam cycle)	21,756	21,455	25,159	27,651
Performance Without Let Down Turbine				
Net Plant Output - kw	607,211	590,335	673,013	777,065
Overall Plant Efficiency, %	25.9	25.1	26.7	30.8
Performance With Let Down Turbine				
Net Output From Let Down Turbine - kw	0	43,329	27,736	166,583
Net Change in Steam Cycle Output - kw	-11,158	-28,308	-23,183	-81,992
Net Plant Output - kw	596,053	605,356	677,566	861,656
Overall Plant Efficiency, %	25.4	25.8	26.9	34.2

*Numbers in parenthesis indicate Latent Heat from Condensation of Gasifier Steam

**Numbers in parenthesis indicate Fuel Energy in MM Btu Needed to Produce Equivalent Power

Gasifier Performance - From the process data for each system, the total energy in the clean product gas can be obtained. The energy in the gas used in sulfur recovery must be added to this to give the actual clean fuel gas energy leaving the gasifier. In addition, the gasifier must be credited with heat extracted from the gasifier and fuel stream that is used elsewhere in the system and also must be debited for heat that is input to the gasifier and fuel stream. The quantity of heat rejected to the cooling tower or to air coolers is indicated. It represents a thermal loss similar to that due to radiation or slag losses from the gasifier. In all cases, gasifier steam and air preheat requirements represent a significant fraction of the input coal energy. The net heat required by the gasifier is then divided by boiler efficiency to give an effective parasitic fuel loss. This, subtracted from the actual clean fuel gas energy leaving the gasifier, gives the net fuel energy out of the gasifier.

In all cases, it can be seen that the sum of actual clean fuel gas energy leaving the gasifier and rejected from the gasifier and fuel gas stream do not equal the sum of the coal energy input and net heat input. The difference is made up of heat losses from the equipment, slag, latent heat of steam in the product gas, heat rejected from the product gas, and a small loss from the sulfur which is not credited to the fuel gas.

The relatively poor performance of the Bu Mines/Selexol combination is due to the need to cool the gas prior to cleanup. This cooling must be done with a water quench, since the gas contains tars that would foul heat exchange surfaces. More of the heat could be removed by condensing the water vapor in the gas stream following the quench and tar removal step; however, that is at a low temperature and it would be difficult to justify economically.

In the Bu/Mines Selexol system all of the heat extracted from the gas stream involves condensation of water vapor. Because all of the steam injected into the gasifier is not used in the reaction, about 72 percent of the steam input to the Bu Mines gasifier (vs. 63 percent for the BCR gasifier) is present in the gasifier exit stream. The resultant dew point is on the order of 250 to 280 F. As the fuel gas stream temperature is dropped below this point, some of the latent heat of that steam is either utilized or rejected. To indicate the extent to which that happens, the values in Table 43 shown in parenthesis represent that part of the heat that is removed by condensation of the gasifier steam. While it may be argued that the effects of cooling the gasifier effluent for use with a low-temperature cleanup should not be charged to the gasifier, it is interesting to note that in the BCR/Selexol System the need for cooling is turned into an advantage by making use of part of the latent heat of the gasifier steam. Because of this ability to take advantage of the latent heat, and also because the process heat requirements for the

BCR/Consol system must be met by burning fuel at boiler efficiency, the net fuel out of the gasifier is slightly higher for the BCR/Selexol System than for the BCR/Consol. Note that while 290 MM Btu/hr is rejected to cooling water or to air from the BCR/Selexol System, it does not represent a major penalty since a large part of that is latent heat of water vapor that would be otherwise lost through the stack, were it not condensed here.

It is possible that further use could be found for the low-temperature heat resulting from the water quench following the Bu Mines gasifier. While the effective gasifier would be improved, the ability to use other low-temperature heat would certainly be impaired and a significant change in overall performance would not be expected. As all of these systems show relatively poor performance potential, such improvements were not investigated. The fuel energy equivalent of auxiliary power can be obtained by dividing by boiler and steam cycle efficiency. These numbers are shown in parenthesis in the auxiliary power tabulation. When charged to the appropriate part of the system, this gives a measure of overall performance. For the gasifiers, auxiliary power requirements reduce the Bu Mines performance by 1.2 points and the BCR by 0.37 points. Coal preparation is the largest requirement for auxiliary power.

Cleanup System Performance - Sulfur and ammonia removal have been treated separately to provide a view of the relative penalty associated with the removal of each and to permit a better comparison of high- and low-temperature systems, since the high-temperature systems have no provision for ammonia removal. In both low-temperature systems, ammonia accounts for a significant part of the process heat used. When sulfur removal is considered, the Bu Mines/Selexol System compares quite favorably to the BCR/Consol performance, requiring less process heat (110 vs. 122 MM Btu/hr), but more power (14,350 vs. 6,381 kw). The BCR/Selexol System performance is degraded due to the relatively high concentrations of COS that must be removed. The Selexol solvent has a lower capacity for COS and the solvent flow rate must be increased to provide high COS removal efficiency. Thus, increases are required in power for pumping and refrigeration and in heat for regeneration. In addition, the high solvent flow rates result in a larger amount of CO_2 absorption which in turn, lowers the concentration of H_2S in the feed to the Claus plant. This causes an increase in complexity and fuel requirements. A possible alternative would be to design the cleanup system for H_2S removal only. This would result in about 800 ppm of sulfur compounds in the product gas and would reduce utility requirements to a level comparable to the Bu Mines/Selexol system.

The high-temperature sintered iron oxide cleanup process shows up poorly in this comparison primarily because of the energy needed to obtain elemental sulfur from the SO_2 that is produced upon regeneration of the iron oxide sorbent. The production of SO_2 results in a net export of heat from the sulfur absorption process and virtually all of the utility requirements for this cleanup system are associated with the subsequent recovery of elemental sulfur. The feed gas to the recovery unit is dilute and requires concentration followed by reduction of a portion of the SO_2 to form the H_2S needed in a Claus reaction to form elemental sulfur. The system also has a relatively high auxiliary power requirement about 2/3 of the power required by the sulfur recovery section. Another means of disposing of the SO_2 desorbed from the iron oxide would be necessary to make this system attractive.

The heat available to the steam cycle is the net fuel energy out of the cleanup process multiplied by the boiler efficiency. The resultant overall efficiency from input coal energy to boiler output tends to favor the two BCR systems. However, the high auxillary power requirements for the BCR/Selexol combination negate that advantage and the net overall efficiency is about equal for all except the BCR/Consol system which benefits from a high-temperature cleanup system with very low auxiliary power and process heat requirements.

Steam Cycle Performance - The steam cycle output power neglects the power that is required to compress the supply air to the gasifier and the power that is available from the expansion of the product gas from gasifier pressure to burner inlet. Therefore, a compressor and turbine must be superimposed on the system to determine overall performance.

The approach used in estimating performance with a let down turbine was to approximate turbine power available in an expansion from 500 psia. It was assumed to be directly proportional to product gas flow rate and absolute fuel gas temperature. Turbine output power is used to drive the gasifier air supply compressor with the excess power available for electric generation. Required compressor prower was taken to be directly proportional to gasifier inlet air flow. The net output from the letdown turbine is given in Table 43 and includes an allowance for plant auxiliary power. Where turbine power was less than that needed to drive the compressor, as in the Bu Mines/Selexol system, it was assumed that the fuel temperature could be increased by regeneration or by partial combustion to a point where the turbo compressor was self-sustaining. For the Bu Mines/Selexol system, that temperature is approximately 760 F.

The effect of the letdown turbine on steam cycle output was estimated by assuming that 83 percent of the compressor work was available as thermal energy and could be used to raise process steam as well as preheat the air to the gasifier. The remainder is wasted in aftercooling between stages to minimize compressor power and limit the exit temperature to a level compatible with common blade materials. All of the turbine power is subtracted from the heat available to the steam cycle. The net change in plant output was then calculated based on steam cycle heat rate and correcting for the appropriate auxiliaries.

The effect of product gas temperature is emphasized by the addition of the letdown turbine. The 1700 F gas temperature out of the Consol cleanup system results in an increase of approximately 10 percent in net plant output.

To evaluate the effect of the approximations used in the foregoing analysis and to provide a reference case for a gasified coal - steam cycle system, the BCR/Consol/steam system shown schematically in Figure 54 was analyzed using the SOAPP preprocessing system. (Appendix A). This is basically a system bookkeeping tool that allows the user to assemble a power system using a number of standard modules and provides the computation logic necessary to determine performance of the system. For the BCR/Consol/steam cycle system, the computer configuration was that shown in Figure 55. The main stream includes provision for process steam generation in a low-pressure boiler as well as a deaerator for boilers external to the main boiler and steam cycle. The main boiler and steam cycle combines a combustor operating at 10 percent excess air with a steam boiler and steam cycle having the performance characteristics previously shown in Figure 13. Stack temperature is a minimum of 300 F.

The gasification stream consists of three stages of compression with intercooling and heat recovery where possible. A side stream contains the CO_2 separation module needed to provide CO_2 for dolomite regeneration with the remainder of the gas used to transport coal into the gasifier.

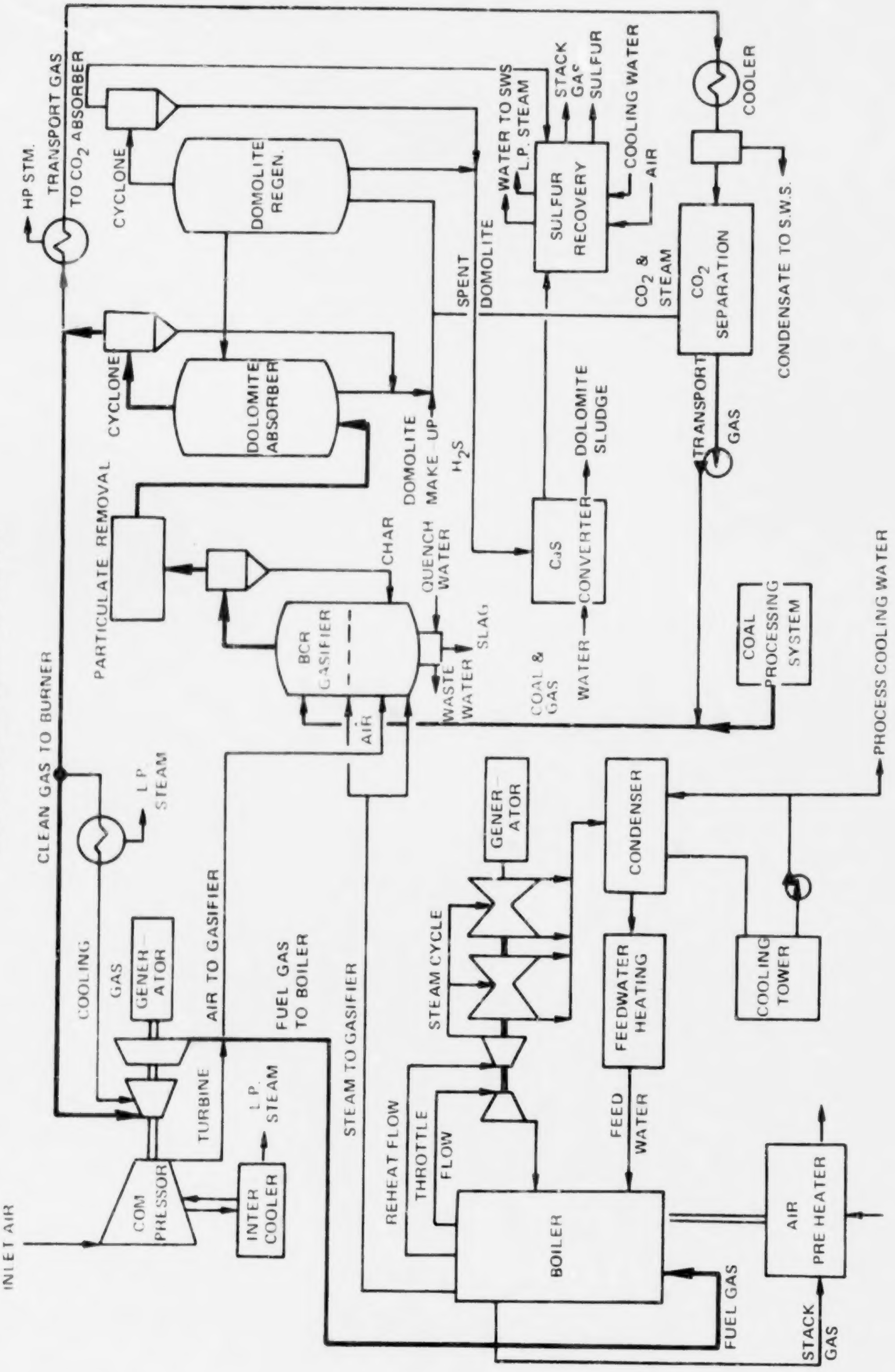
Using this model, the performance characteristics presented in Table 44 were calculated. They differ from the results shown in Table 43 by about 0.5 points in overall efficiency. This difference is due to the need for cooling gas in the tubine. At the 1700 F turbine inlet temperature, cooling would be required to keep blade temperatures to a level compatible with long life, e.g., 1500 F. Air cannot be bled from the compressor to be used since there would be the danger of local combustion at the coolant discharges in the blades. Therefore, it was necessary to devise a cycle in which a part of the turbine inlet gas (8.7 percent) is cooled and used instead of compressor bleed air for engine cooling.

COMBINED-CYCLE SYSTEMS

A summary of the calculated performance characteristics for each of the four systems is given in Table 45. The performance of the second-generation gasification and cleanup systems mated with a first-generation power system is

FIG. 54

BCR/CONSOL/STEAM SYSTEM


INLET AIR
COM PRESSOR
INTER COOLER
L.P. STEAM
TURBINE
GAS
GENER ATOR
COOLING
CLEAN GAS TO BURNER
L.P. STEAM
AIR TO GASIFIER
FUEL GAS TO BOILER
STEAM TO GASIFIER
AIR
BCR GASIFIER
WASTE WATER
SLAG
QUENCH WATER
CHAR
PARTICULATE REMOVAL
DOLOMITE ABSORBER
CYCLONE
DOMOLITE MAKE-UP
H_2S
HP STM.
TRANSPORT GAS TO CO_2 ABSORBER
CYCLONE
DOMOLITE REGEN.
SPENT DOMOLITE
CaS CONVERTER
WATER
COAL & GAS
DOLOMITE SLUDGE
SULFUR RECOVERY
WATER TO SWS
L.P. STEAM
STACK GAS
SULFUR
COOLING WATER
AIR
CO_2 & STEAM
CO_2 SEPARATION
CONDENSATE TO S.W.S.
COOLER
TRANSPORT GAS
COAL PROCESSING SYSTEM
PROCESS COOLING WATER
REHEAT FLOW
THROTTLE FLOW
STEAM CYCLE
GENER ATOR
CONDENSER
FEEDWATER HEATING
FEED WATER
COOLING TOWER
BOILER
FUEL GAS
AIR PRE HEATER
STACK GAS

RO2-187-2

FIG. 55

GASIFIED COAL–STEAM SYSTEM

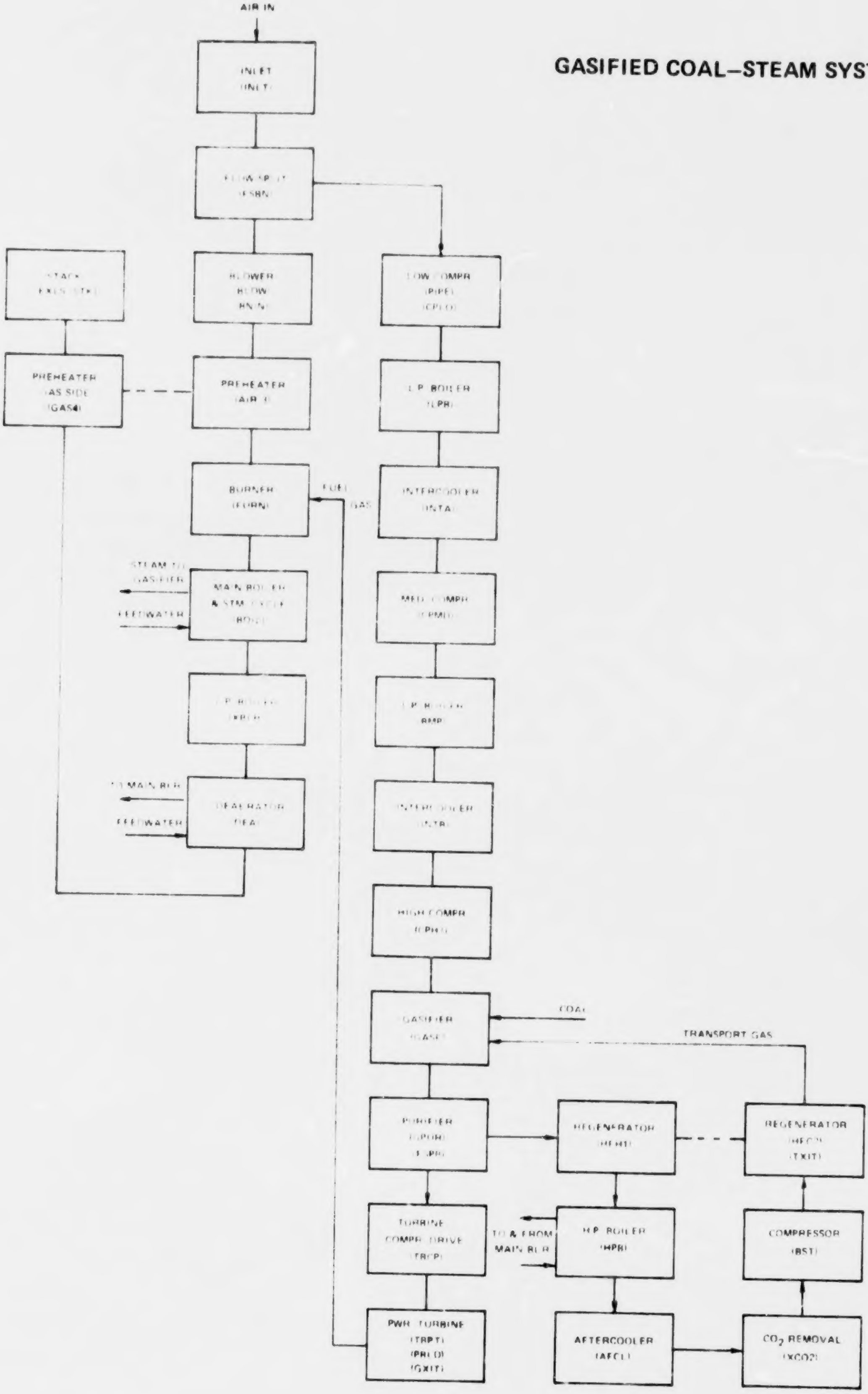

AIR IN
INLET (INLT)
FLOW SPLIT (FSBN)
STACK EXIT (STK)
BLOWER (BLOW) (BNIN)
LOW COMPR (PIPE) (CPLO)
PREHEATER GAS SIDE (GAS4)
PREHEATER (AIR 3)
L.P. BOILER (LPB)
BURNER (FURN)
FUEL GAS
INTERCOOLER (INTA)
STEAM TO GASIFIER
FEEDWATER
MAIN BOILER & STM CYCLE (BOIL)
MED COMPR (CPMD)
L.P. BOILER (XBLR)
I.P. BOILER (BMP)
TO MAIN BLR
FEEDWATER
DEAERATOR (DEA)
INTERCOOLER (INTB)
HIGH COMPR (CPHI)
GASIFIER (GASF)
COAL
TRANSPORT GAS
PURIFIER (PURI)
REGENERATOR (REHT)
REGENERATOR (REC2) (TXIT)
TURBINE COMPR DRIVE (TRCP)
TO & FROM MAIN BLR
H.P. BOILER (HPB)
COMPRESSOR (BST)
PWR TURBINE (TRPT) (PRLD) (GXIT)
AFTERCOOLER (AFCL)
CO2 REMOVAL (XCO2)

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Table 44

BCR/Consol/Steam Cycle Performance

Fuel System	
Coal Feed Rate - lb/hr	700,000
Gasifier Exit Temp - F	1,800
Cleanup System Exit Temp - F	1,700
Fuel Gas HHV - Btu/scf	135.8
Boiler & Steam Cycle	
Fuel - Air Ratio	.7195
Air Preheat Temp - F	376
Throttle Press - psia	2,400
Throttle & Reheat Temp - F	1,000
Net Heat Rate - Btu/kwh	7,836
Net Stm Cycle Output - Mw	760.5
Let Down Turbine	
Net Power Output - Mw	143.2
Auxiliary Power	
Gasifier & Cleanup - Mw	27.8
Steam Cycle - Mw	25.2
Gas Turbine - Mw	.8
Plant Performance	
Net Power Output - Mw	849.9
Efficiency (HHV)	33.7

Table 45

INTEGRATED SYSTEMS PERFORMANCE SUMMARY

	First Generation		Second Generation		Second Generation Gasification First Generation Power System	
	BOM/ Selexol	BOM/ Iron Oxide	BCR/ Selexol	BCR/ Consol	BCR/ Selexol	BCR/ Consol
Gas Turbine						
Turbine Inlet Temperature - F	2,200	2,200	2,600	2,600	2,200	2,200
Compressor Pressure Ratio	16	16	24	24	16	16
Exhaust Temperature - F	916	913	1,107	1,115	913	920
Output Power - Mw	595.4	626.2	726.6	857.6	642.3	757.6
Steam Cycle						
Steam Temperature - F	816	813	1,000	1,000	813	820
Steam Pressure - psia	1,250	1,250	1,250	1,250	1,250	1,250
Condenser Pressure In. Hg. Abs.	4.0	4.0	4.0	4.0	4.0	4.0
Net Steam Cycle Output - Mw	223.8	208.1	293.3	296.6	273.5	271.4
Net Steam Cycle Efficiency	.280	.292	.307	.307	.282	.279
Gasifier and Cleanup System						
Coal Feed Rate - lb/hr.	700,000	700,000	700,000	700,000	700,000	700,000
Air - Coal Ratio	3.113	2.688	3.088	3.088	3.088	3.088
Steam - Coal Ratio	.405	.349	.567	.567	.567	.567
Air Preheat Temperature - F	800	800	800	800	800	800
Steam Temperature - F	564	564	1,000	1,000	913	920
Steam Pressure - psia	1,250	1,250	1,250	1,250	1,250	1,250
Gasifier Exit Temperature - F	1,000	1,000	1,800	1,800	1,800	1,800
Cleanup System Exit Temperature - F	265	1,070	1,000	1,700	1,000	1,700
Fuel Gas Higher Heating Value Btu/SCF	160.3	169.9	159.3	135.8	159.3	135.8
Integrated System						
Gross Power - Mw	819.2	834.3	1,019.9	1,154.2	915.8	1,029
Boost Compressor Power - Mw	43.4	36.1	40.1	40.2	40.1	40.2
Gasifier & Cleanup Aux. Power - Mw	28.2	36.5	58.7	27.6	58.7	27.6
Plant Auxiliaries - Mw	10.6	10.2	13.6	14.5	12.5	13.1
Net Plant Output - Mw	737.0	751.5	907.5	1071.9	804.5	948.1
Net Plant Efficiency (HHV-Coal)	.314	.320	.360	.425	.319	.376

also given to permit a gross comparison of the gasification and cleanup systems. Looking at overall plant efficiency, it is clear that there are two significant factors in performance improvement: turbine inlet temperature and a high-temperature cleanup system. A closer look will show that the benefits of the high-temperature system are due to a highly efficient process (lower requirements for process steam and auxiliary power) as well as the ability to operate at high-temperature. Unfortunately, the performance gain with the high-temperature system is at the expense of ammonia and particulate removal capability.

System Descriptions

The general arrangements for each of the four integrated systems are presented here. Performance characteristics and detailed utility requirements are given in the following sections.

Bureau of Mines/Selexol - This system is shown schematically in Figure 56. The gasifier is supplied with air and steam from the COGAS system. Compressor bleed air is used to generate process steam prior to entering a boost compressor. Saturated steam is extracted from the high-pressure section of the waste-heat boiler. Because of the tar content, the gasifier exit stream is quenched and the tars separated and returned to the gasifier. The gas leaving the water quench is saturated with water vapor at 338 F and as a result, little regeneration of the fuel gas is possible. However, the latent heat of the water vapor is quite significant and it is used in the reboiler of the sour water stripper. After regeneration with the product gas, the remaining heat is rejected to ambient to lower the temperature prior to a water scrub and Selexol cleanup tower. Condensate from the Selexol stripper and the gas cooling process along with water from the scrubber must be processed in the sour water stripper. In the ammonia recovery process, H_2S absorbed in the water is recovered and sent to the Claus plant. The requirements for each of these processes for heat, fuel gas or power have been previously discussed.

After regeneration, the clean fuel gas is fed to the COGAS system which operates in a conventional manner with the exception that steam is extracted from the high-pressure section of the boiler for subsequent use in the gasifier and the ammonia recovery unit. Also, steam raised in the gasifier jacket is used to supplement the waste-heat boiler. Because the fuel gas stream provides heat directly to the sour water stripper, the quantity of low pressure steam required is relatively small and the steam raised in the gasifier bleed air stream is sufficient to satisfy this need.

BUMINES SELEXOL SYSTEM

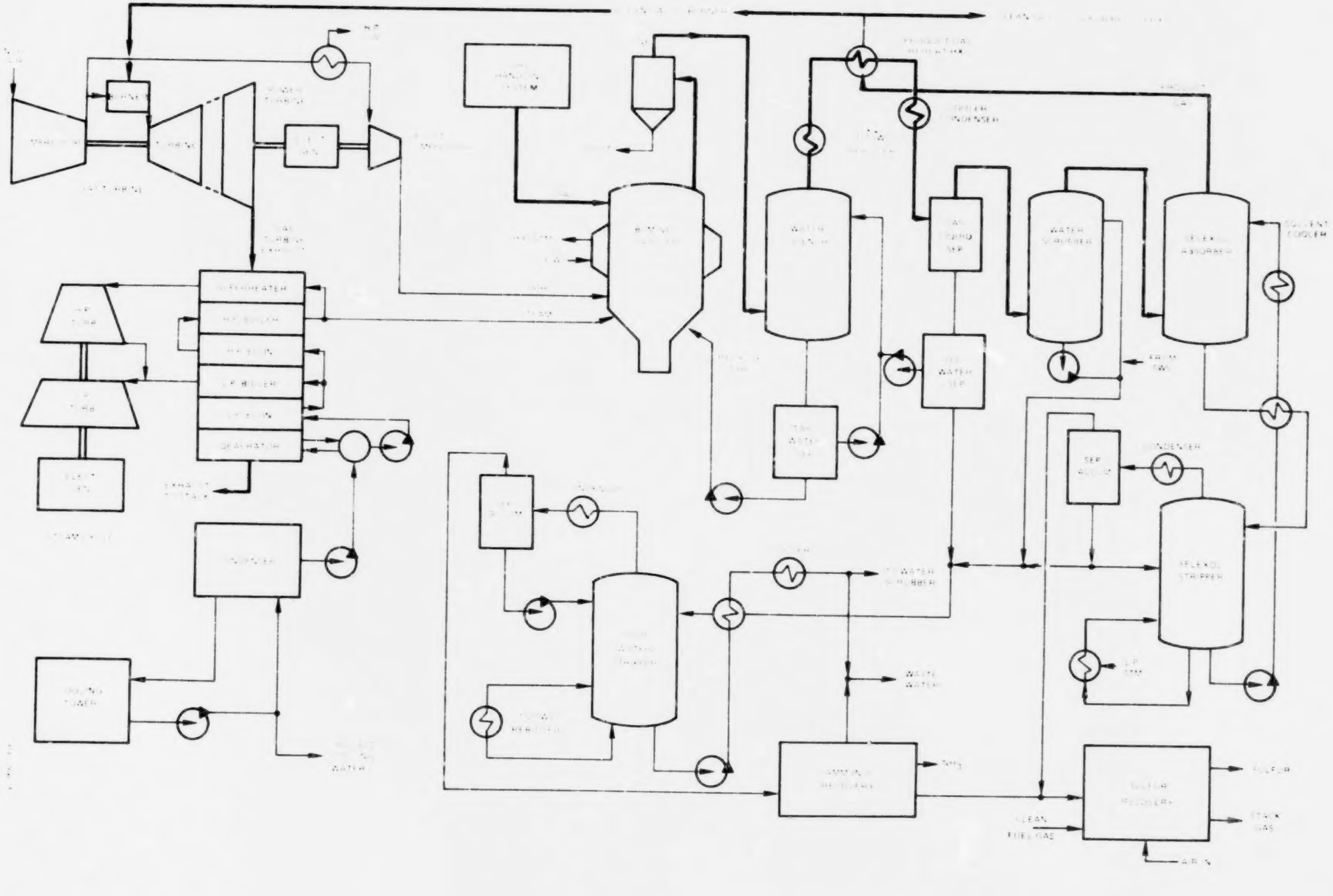

BURNER
TURBINE
POWER TURBINE
ELECT GEN
GAS TURBINE EXHAUST
SUPERHEATER
H.P. BOILER
H.P. ECON
L.P. BOILER
L.P. ECON
DEAERATOR
H.P. TURB
L.P. TURB
ELECT GEN
EXHAUST TO STACK
STEAM CYCLE
CONDENSER
COOLING TOWER
WATER
COAL HANDLING SYSTEM
AIR
STEAM
WATER QUENCH
PRODUCT GAS REHEATER
BOILER CONDENSER
GAS LIQUID SEP
OIL WATER SEP
TAR WATER TANK
WATER SCRUBBER
SELEXOL ABSORBER
PRODUCT GAS
SOLVENT COOLER
FROM SWS
CONDENSER
SEP ACCUM
SELEXOL STRIPPER
L.P. STM
TO WATER SCRUBBER
WASTE WATER
AMMONIA RECOVERY
NH_3
SULFUR RECOVERY
SULFUR
STACK GAS
CLEAN FUEL GAS
AIR IN

FIG.

Bureau of Mines/Sintered Iron Oxide - Because the tar in the product gas does not affect operation of the cleanup system, the water quench can be eliminated with this high-temperature cleanup system. The flow schematic, shown in Figure 57, is quite simple due to the lack of regeneration, and the elimination of the water scrub and tar recycle. Gasifier performance is enhanced by not recycling the tar since it permits operation at lower air- and steam-to-fuel ratios. The net cleanup system reactions are exothermic both during absorption and desorption adding sensible energy to the fuel and enabling the recovery of heat in the form of process steam from the regenerator off-gas. However, the sulfur comes off the iron oxide as SO_2 and in the sulfur recovery system must be concentrated and a portion reduced to H_2S so that it can be used to form elemental sulfur in a Claus unit. This process uses a considerable amount of fuel and process steam. As a result, even with the steam raised in the process, a significant amount of low-pressure steam must be taken from the waste heat boiler to satisfy these requirements. The lower gasifier steam requirement means that nearly all the heat needed to raise the gasifier steam can be taken from the gasifier jacket cooling.

The system suffers from the inability to remove ammonia from the gas and the uncertainties associated with particulate removal. Also, the tars will contain some nitrogen and sulfur compounds which will not be touched by the cleanup system.

BCR/Selexol System - With this system, shown in Figure 58, it is possible to both regenerate and extract useful heat from the gasifier exit stream. The high gas turbine exhaust temperature resulting from the 2600 F turbine inlet temperature allows sufficient superheat in the steam to satisfy the gasifier heat balance without using the transport gas regenerator that has been included in the computer model and used in the parametric analyses. Ammonia and H_2S removal are quite similar to the Bu Mines system, but the Selexol System is complicated by a high percentage of COS in the gas to be processed. Because COS is about one-third as soluble as is the H_2S, its presence in gas streams requiring sulfur limitations lower than that accomplished by a 30 to 40 percent COS removal, require that the system be designed specifically for COS. The utilities for this system reflect this situation. A significant amount of heat is recovered from the gasifier jacket and the high temperature gasifier exit stream. This heat is used to raise high pressure steam which supplements the high pressure section of the waste heat boiler by adding about 50 percent to the steam raised there. After the steam for the gasifier is extracted, the net increase in steam available for the turbine is about 25 percent. All of the low-pressure process steam requirements are met by steam raised in the boilers located upstream of the boost compressor and downstream of the fuel gas regenerator.

BUMINES SINTERED IRON OXIDE SYSTEM

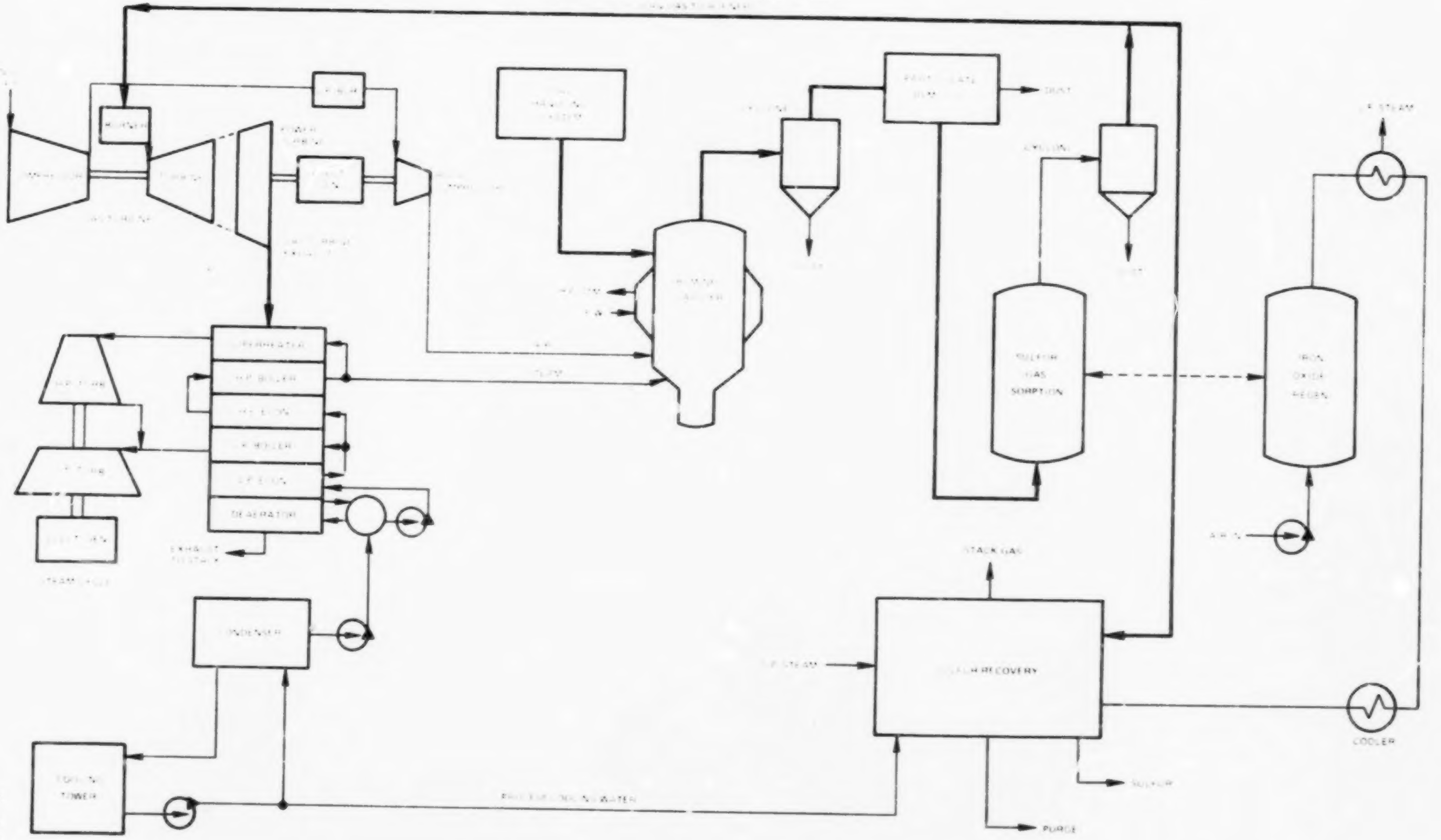
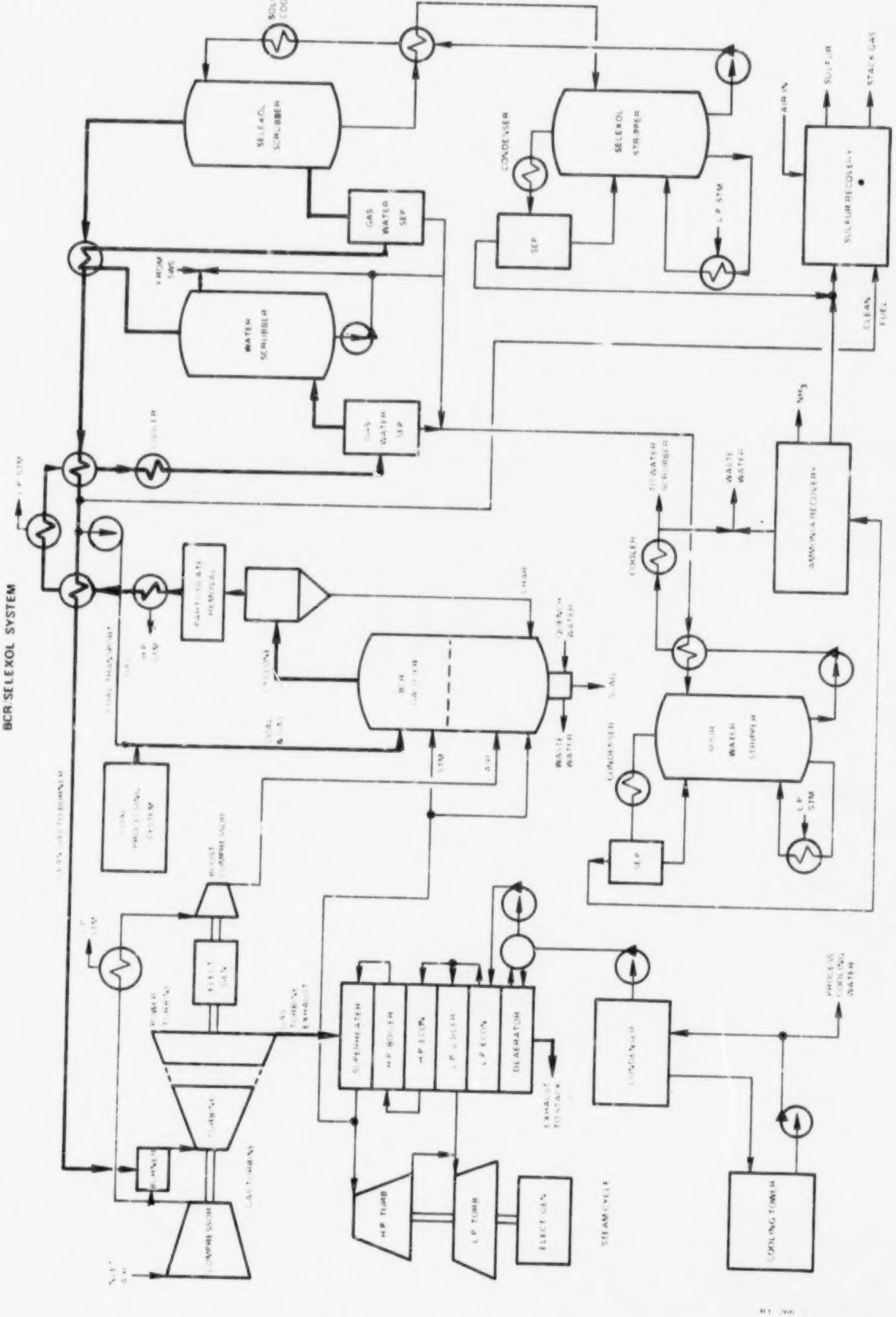

BURNER
POWER TURBINE
SUPERHEATER
H.P. BOILER
H.P. ECON
L.P. BOILER
L.P. ECON
DEAERATOR
EXHAUST TO STACK
STEAM CYCLE
CONDENSER
CYCLONE
PARTICULATE REMOVAL
DUST
CYCLONE
DUST
SULFUR GAS SORPTION
IRON OXIDE REGEN
L.P. STEAM
AIR IN
STACK GAS
SULFUR RECOVERY
COOLER
SULFUR
PURGE
PROCESS COOLING WATER
COOLING TOWER

FIG. 57

FIG. 58


BCR/SELEXOL SYSTEM

SOLVENT COOLER
SELEXOL SCRUBBER
SELEXOL STRIPPER
CONDENSER
SEP
L.P. STM
SULFUR RECOVERY
AIR IN
SULFUR
STACK GAS
CLEAN FUEL
GAS WATER SEP
FROM SWS
WATER SCRUBBER
NH3
AMMONIA RECOVERY
WASTE WATER
TO WATER SCRUBBER
COOLER
SOUR WATER STRIPPER
CONDENSER
SEP
L.P. STM
L.P. STM
BOILER
GAS WATER SEP
PARTICULATE REMOVAL
CHAR
QUENCH WATER
SLAG
CYCLONE
BCR GASIFIER
WASTE WATER
STM
AIR
COAL & CHAR
H.P. STM
COAL TRANSPORT GAS
COAL PROCESSING SYSTEM
BOOST COMPRESSOR
CLEAN GAS TO BURNER
L.P. STM
ELECT GEN
POWER TURBINE
GAS TURBINE EXHAUST
SUPERHEATER
H.P. BOILER
H.P. ECON
L.P. BOILER
L.P. ECON
DEAERATOR
EXHAUST TO STACK
CONDENSER
PROCESS COOLING WATER
COOLING TOWER
STEAM CYCLE
H.P. TURB
L.P. TURB
ELECT GEN
BURNER
TURBINE
COMPRESSOR
GAS TURBINE
INLET AIR

BCR/Consol System - In this system, shown in Figure 59, the gas steam arrangement is very simple and the omission of the aqueous cleanup system and its associated stripping unit significantly reduces process steam requirements. As a result, it is possible to raise high-pressure steam in boilers located upstream of the boost compressor and in the transport gas stream prior to the CO_2 removal unit. The low-pressure process steam needed to supplement that raised within the cleanup and sulfur recovery system can be supplied by a boiler located between the high-pressure boiler and boost compressor. The high-pressure steam is used to supplement the waste heat boiler and balances the amount of superheated steam extracted for injection into the gasifier. As in the case of the BCR/Selexol system, steam superheat is sufficient to satisfy the gasifier heat balance so that the transport gas regenerator was not needed.

Performance

The SOAPP computer routine was used to calculate the performance characteristics of the four selected systems. The configuration shown in Figure 30 was used for all but the BCR/Consol system which was represented by the configuration of Figure 31. Using the SOAPP routine with the final system configurations, it was possible to re-evaluate some of the parametric integration work discussed in an earlier section. It was found that the basic results of that work were valid. As an example, the effects of pressure ratio and turbine inlet temperature on the BCR/Selexol system are presented in Figure 60 and show the same characteristics as those previously presented, thereby assuring that the selection of operating pressure was valid. It should be noted that these estimates were limited to the effects of changes in the COGAS system and associated compressor power variations because of the fixed input-output model used for the gasifier and cleanup systems.

The COGAS system performance as a function of the gas turbine pressure ratio for both first- and second-generation turbine systems is shown in Figures 61 and 62. To provide consistency, both figures include the performance of distillate fired systems which can be compared to those using the gas generated in either a BCR/Selexol or BCR/Consol gasifier-cleanup system. From these figures it is clear that the key to improved system performance is in the gasifier/cleanup performance. Therefore, it is necessary to understand the effect of individual component performance in terms of its net effect on the overall system. This relationship is quite complex due to the interdependency of the various parts of the COGAS system. An accounting system must take into consideration the difference between fuel energy and thermal energy or steam. Whereas the fuel can be used at combined cycle efficiency (about 50 percent), the thermal energy or steam can be used only at steam cycle efficiency (about 30 percent). Therefore, to make a proper accounting of each

BCR/CONOCO SYSTEM

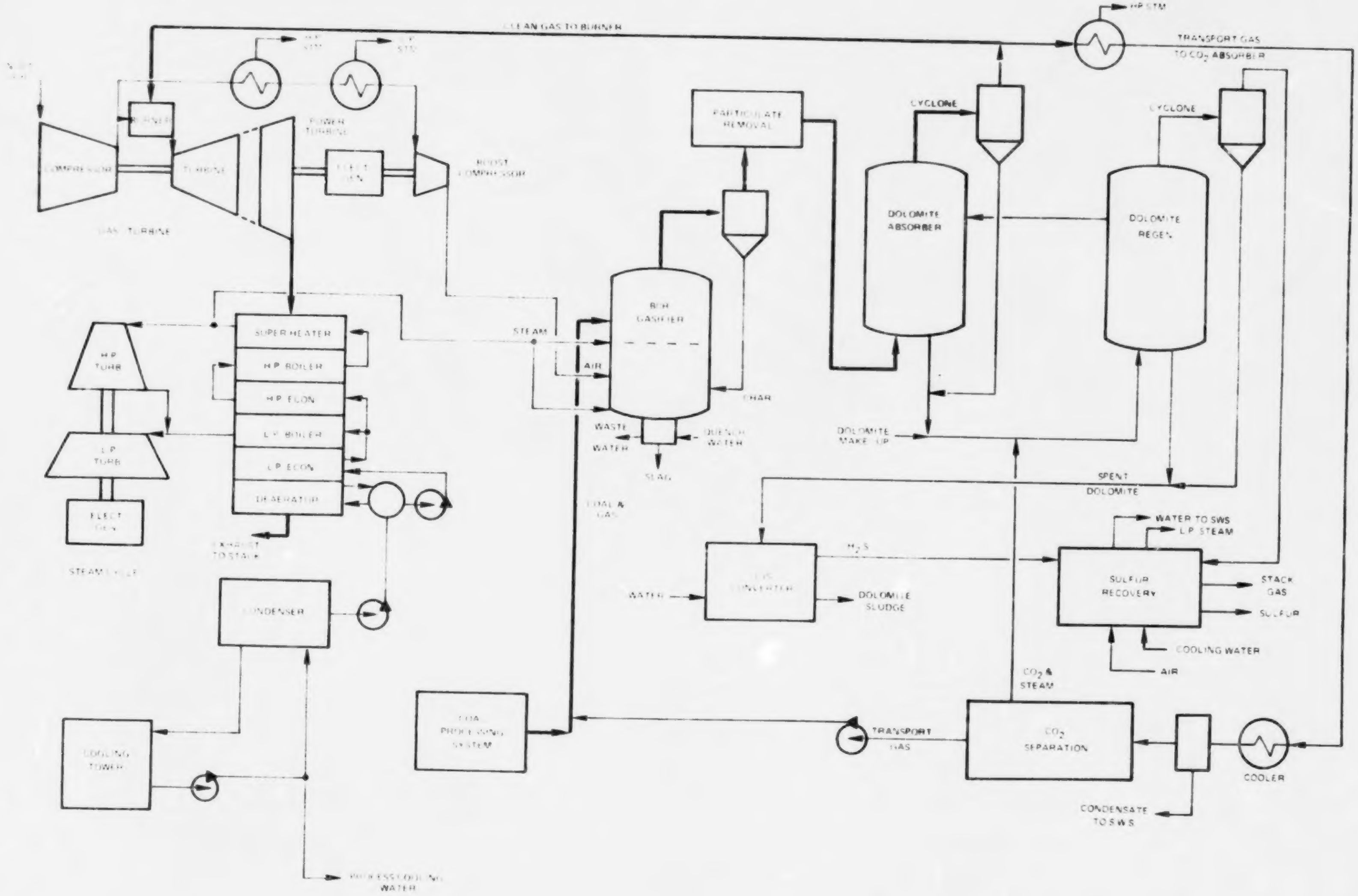

CLEAN GAS TO BURNER
H.P. STM
L.P. STM
HP STM
TRANSPORT GAS TO CO_2 ABSORBER
BURNER
COMPRESSOR
TURBINE
POWER TURBINE
ELECT GEN
BOOST COMPRESSOR
GAS TURBINE
PARTICULATE REMOVAL
CYCLONE
CYCLONE
DOLOMITE ABSORBER
DOLOMITE REGEN
BCR GASIFIER
STEAM
AIR
CHAR
WASTE WATER
QUENCH WATER
SLAG
COAL & GAS
DOLOMITE MAKE UP
SPENT DOLOMITE
SUPER HEATER
H.P. BOILER
H.P. ECON
L.P. BOILER
L.P. ECON
DEAERATOR
EXHAUST TO STACK
H.P. TURB
L.P. TURB
ELECT GEN
STEAM CYCLE
CONDENSER
COOLING TOWER
PROCESS COOLING WATER
COAL PROCESSING SYSTEM
CaS CONVERTER
WATER
H_2S
DOLOMITE SLUDGE
WATER TO SWS
L.P. STEAM
SULFUR RECOVERY
STACK GAS
SULFUR
COOLING WATER
AIR
CO_2 & STEAM
TRANSPORT GAS
CO_2 SEPARATION
COOLER
CONDENSATE TO S.W.S.

FIG. 59

FIG. 60

BCR GASIFIER–SELEXOL CLEANUP PERFORMANCE CHARACTERISTICS

CERAMIC VANES–CONVENTIONAL BLADE COOLING

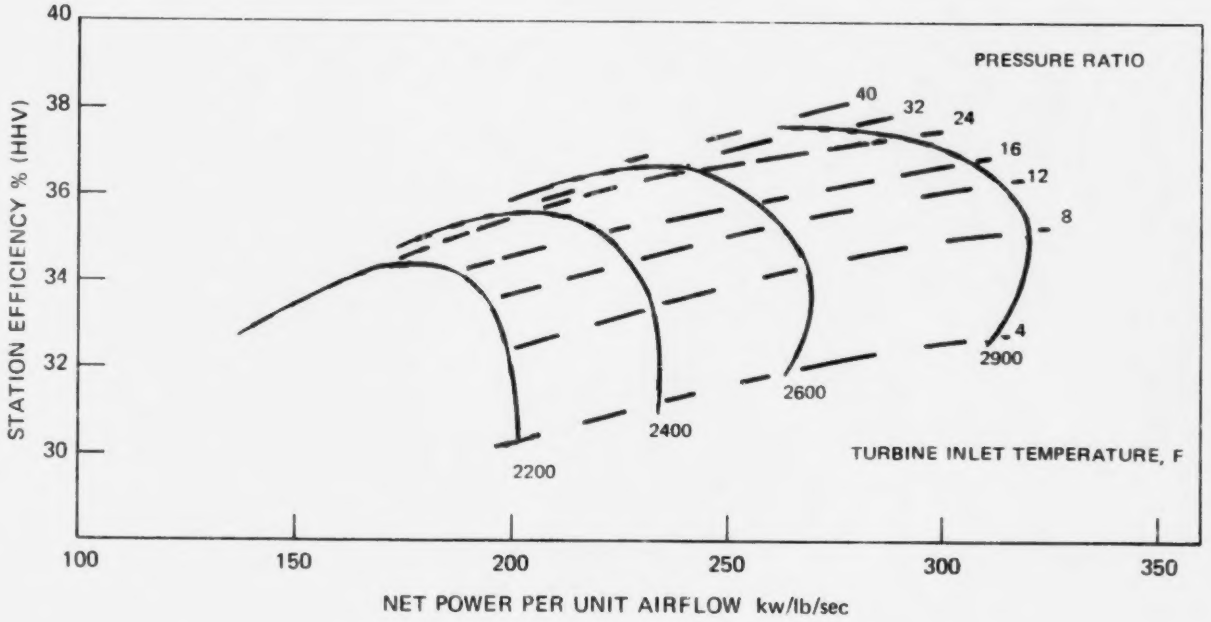


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FIG. 61

PERFORMANCE ESTIMATES FOR SECOND-GENERATION TURBINE SYSTEMS

TURBINE INLET TEMPERATURE = 2600F

CERAMIC VANES, CONVENTIONAL BLADES

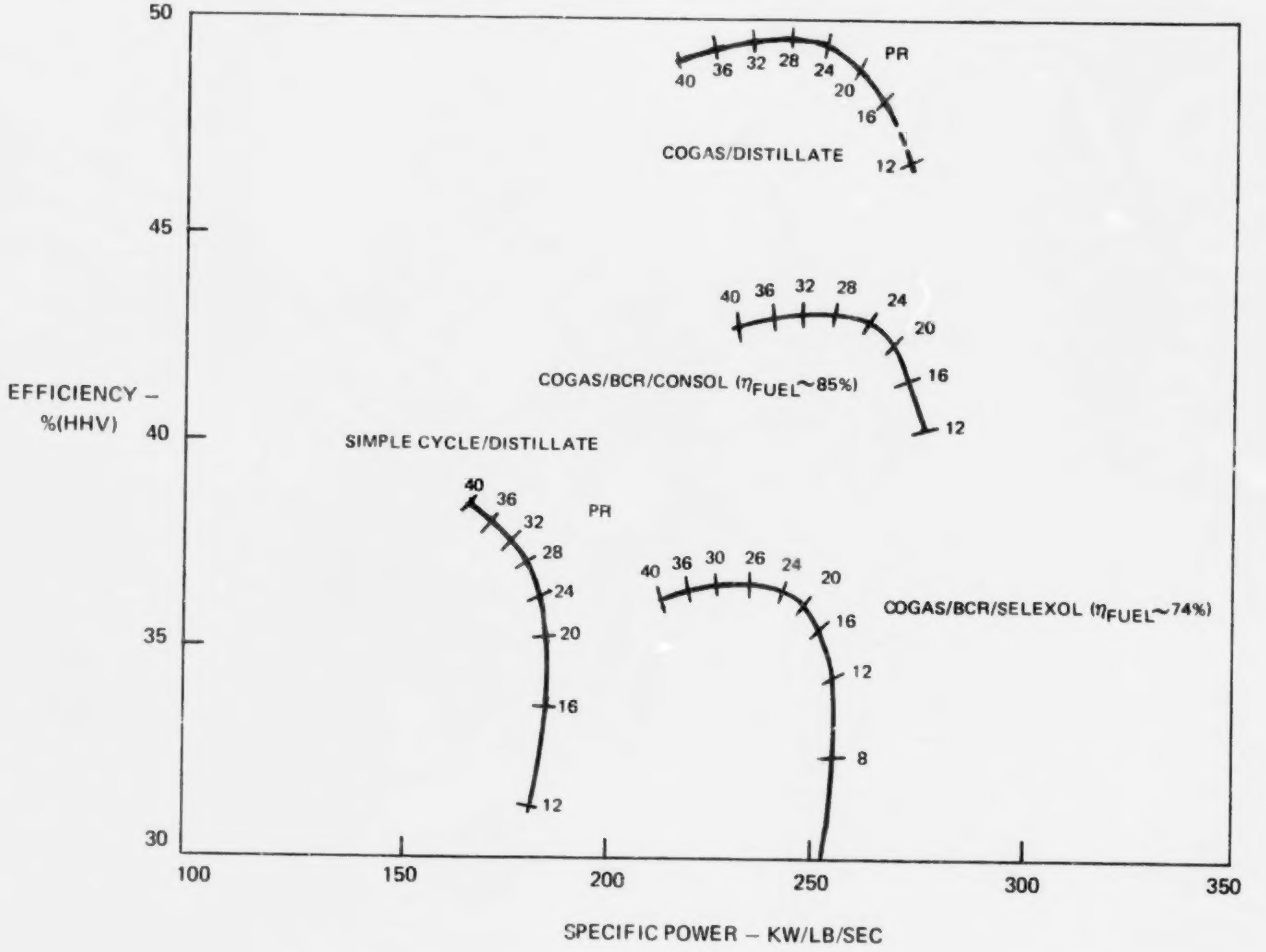
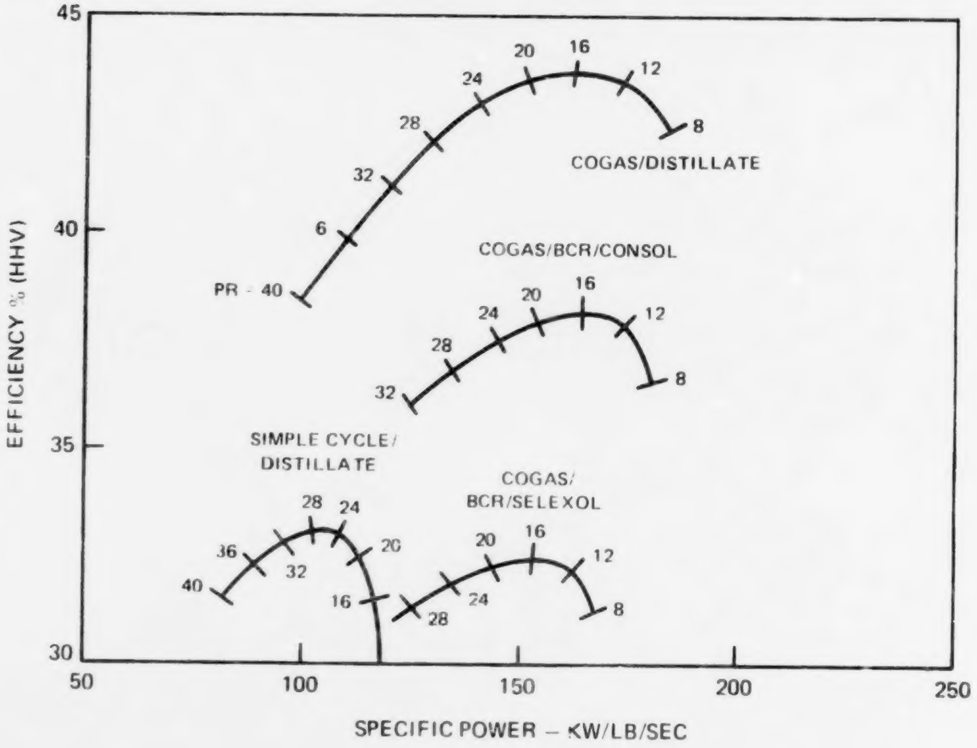


FIG. 62

PERFORMANCE ESTIMATES FOR FIRST GENERATION TURBINE SYSTEMS

TURBINE INLET TEMPERATURE = 2200F

'ONVENTIONAL COOLING



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part of the system it is necessary to separate gas turbine and steam cycle performance characteristics. This has been done for each major system component and the thermal energy gained or lost in each component has been charged at the effective value of fuel energy. Where a component consumes fuel it is charged with the full energy content of that fuel. Process steam or thermal energy consumption is charged only at a rate corresponding to the loss in steam cycle generating capacity. The multiplier used to convert a thermal loss to a fuel energy loss is the ratio of steam to combined cycle efficiency. After all the parasitic losses are subtracted, the output of the gasifier and the cleanup systems is an effective fuel energy which is all used at combined cycle efficiency. In the following paragraphs the accounting system is first described and then applied to the individual systems.

To give a simple mathematical model of the combined-cycle ststem, it is necessary to configure the system as shown in Figure 63. The boost compressor is completely divorced from the gas turbine to keep track of the heat of compression. Because of the general use of higher heating value when dealing with coal, the combined-cycle equations must be altered accordingly. The equations used and their interpretations are given in Appendix E.

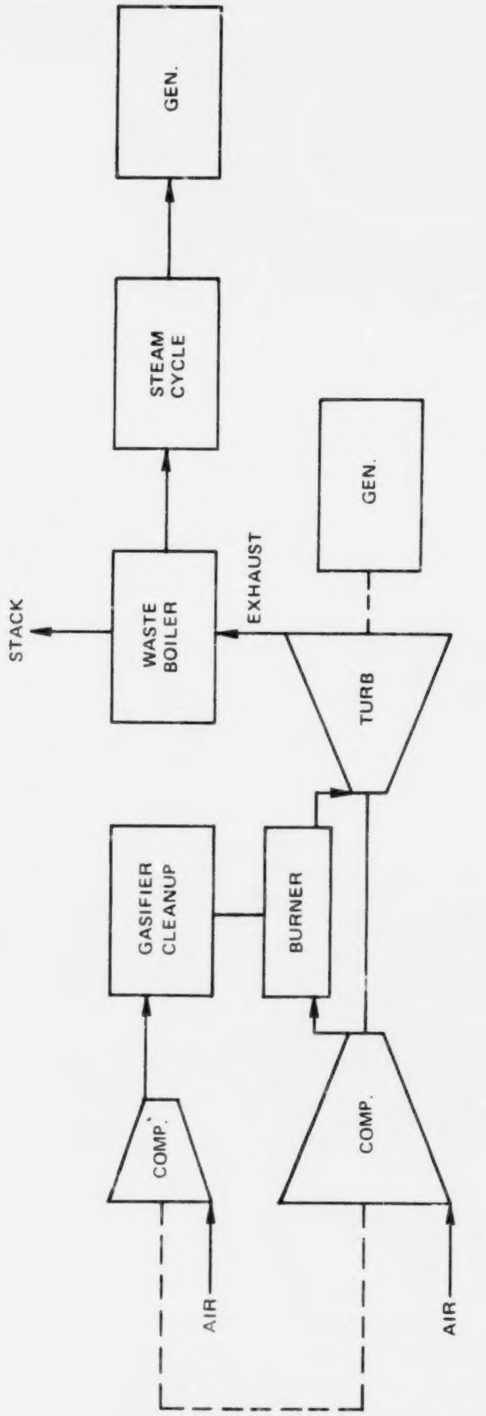
For each of the selected systems, the performance was calculated and the data are presented in Table 46. Gas turbine efficiencies and combined-cycle efficiencies are both higher than the corresponding distillate efficiency. This is due to the high fuel flow rate and the fact that gas turbine output includes the compressor work needed to supply the gasifier air. This will be later subtracted as an auxiliary. The item of interest is the ratio of steam to combined cycle efficiency which is a measure of the effective fuel loss per Btu of process heat required. This is used to relate the value of thermal energy used in process heating to fuel energy, e.g.

$$\text{Thermal Energy } \frac{\eta_{STM}}{\eta_{CC}} = \text{Fuel Energy} \tag{43}$$

This factor has been used to show the effect of the combined cycle energy utilization on each of the four systems in Table 47. Starting with the total energy in the clean fuel gas delivered to the burner, the energy consumed in the cleanup system is added to give a value for energy in the fuel gas from the gasifier but at burner delivery temperature. (For high-temperature cleanup systems this is essentially gasified outlet temperature.) Next the net heat comsumed (or exported) by the gasifier is calculated. This times the ratio of η_{STM}/η_{CC} gives the effective energy removed from (or added to) the fuel stream. The heat rejected from the gas stream is tabulated but does not enter into the calculation. It gives an indication of the effect of the requirement to cool

FIG. 63

COMBINED CYCLE REPRESENTATION


STACK
WASTE BOILER
EXHAUST
STEAM CYCLE
GEN.
GEN.
TURB
GASIFIER CLEANUP
BURNER
COMP.
COMP.
AIR
AIR

R04-48-1

Table 46

COMBINED-CYCLE ENERGY UTILIZATION CHARACTERISTICS

Product Gas	Bu Mines/ Selexol	Bu Mines/ Iron Oxide	BCR/ Selexol	BCR/ Consol
Temperature - F	265	1,070	1,000	1,700
Chemical Energy - MM Btu/hr (HHV)	6,670	6.311	6,564	6,636
Chemical Energy - MM Btu/hr (LHV)	6,208	5,810	6,084	6,133
Sensible Energy (Above 59F) MM Btu/hr	159	778	775	1,729
Total Energy to Burner - MM Btu/hr (HHV)	6,829	7,089	7,339	8,365
Total Energy to Burner - MM Btu/hr (LHV)	6,367	6.588	6,859	7,862
Heating Value Ratio (HVR)	.932	.929	.935	.940
Combined Cycle				
Gas Turbine Output - SHP	953,200	979.880	1.167,924	1,346,582
Gas Turbine Efficiency (η_{GT}) (HHV)	.355	.352	.405	.410
Gas Turbine Outlet Temperature - F	916	913	1,107	1,115
Gas Temperature Out of Boiler - F	300	300	300	300
Boiler Heat Removal Efficiency (Above 59 F) (η_{BLR})	.733	.732	.786	.790
Net Steam Turbine Mech. Output - SHP	297,940	284.920	401,480	406,040
Net Steam Cycle Mech. Efficiency	.286	.298	.313	.314
Combined Cycle Mech. Efficiency (HHV)	.476	.478	.535	.541
Effective Fuel Loss Per Btu Process Heat Required $\left(\frac{\eta_{STM}}{\eta_{CC}}\right)$	.601	.623	.585	.580

Table 47

EFFECT OF COMBINED CYCLE ON GASIFIER & CLEANUP PERFORMANCE

Energy (MM Btu/hr, Unless Otherwise Specified)	Bu. Mines/[illegible]	Bu Mines/Iron Oxide	BCR/Selexol	BCR/[illegible]
Coal Energy Input (HHV)	8,01[illegible]	[illegible],015	8,610	8,610
Fuel To Burner				
Clean Fuel Gas Flow, lb/hr	[illegible]	[illegible]	[illegible],108	3,17[illegible]
Gas Temperature, F	[illegible]	1,070	1,000	1,7[illegible]
Chemical Energy in Gas (HHV)	[illegible]	[illegible]	[illegible]	6,636
Sensible Heat in Gas (above 77 F)	[illegible]	778	775	1,7[illegible]
Total Energy in Gas to Burner (HHV)	[illegible]	7,[illegible]	7,[illegible]	8,3[illegible]
Gasifier				
Clean Fuel Gas Energy Out (at Burner Supply Temperature)	[illegible]	7,[illegible]	7,[illegible]	[illegible]
Heat Available from Gasifier and Gas Stream				
Useful(*)	[illegible]	21[illegible]	1,973 (75)	249
Rejected (*)	574 (181)		293 (155)	65
Process Heat Requirements				
Added to Fuel Gas Stream	[illegible]	74	777	---
Input to Steam	[illegible]	[illegible]	[illegible]	[illegible]
Gasifier Air Preheat	[illegible]	[illegible]	[illegible]	[illegible]
Net Process Heat Used (Required less Useful Heat Available)	[illegible]	[illegible]	[illegible]	[illegible]
Effective Loss in Fuel Energy [illegible]	[illegible]	[illegible]	[illegible]	[illegible]
Effective Energy Out (Clean Gas Energy Less Loss)	[illegible]	[illegible]	[illegible]	8,[illegible]
Effective Gasifier Efficiency	[illegible]	[illegible]	[illegible]	[illegible]
Cleanup System				
Net Process Heat Required in Sulfur Cleanup	[illegible]	[illegible]	[illegible]	[illegible]
Effective Loss in Fuel Energy	[illegible]	[illegible]	[illegible]	[illegible]
Fuel Consumed in Sulfur Recovery	[illegible]	[illegible]	[illegible]	---
Total Effective Fuel Energy for Sulfur Cleanup	[illegible]	[illegible]	[illegible]	[illegible]
Net Process Heat Required in [illegible] Cleanup	[illegible]	---	[illegible]	---
Effective Fuel Energy for [illegible] Cleanup	[illegible]	---	[illegible]	---
Effective Fuel Energy Out of Cleanup	[illegible]	[illegible]	7,[illegible]	7,[illegible]
Effective Fuel Cleanup Efficiency	[illegible]	[illegible]	[illegible]	[illegible]
Combined Cycle Performance				
Heat of Compression [illegible] Air Supply	[illegible]	[illegible]	[illegible]	[illegible]
Effective Fuel Energy [illegible]	[illegible]	[illegible]	[illegible]	[illegible]
Total Effective Fuel Energy to Combined Cycle	[illegible]	[illegible]	7,[illegible]	8,[illegible]
Gross Combined Cycle Shaft Power Output - Mw	[illegible]	[illegible]	[illegible]	1,31[illegible]
Total Compressor Power for [illegible] - Mw	[illegible]	[illegible]	[illegible]	17[illegible]
Gas Turbine Net Electrical Output - Mw	[illegible]	[illegible]	[illegible]	8[illegible]
Steam Turbine Net Electrical Output - Mw	[illegible]	[illegible]	[illegible]	[illegible]
Total Generator Output Power - Mw	[illegible]	[illegible]	[illegible]	[illegible]
Auxiliary Power - kw (% Generator Output)				
Gasifier	[illegible]	[illegible]	[illegible]	21,[illegible] (1.91%)
Gas Cooling	[illegible]			
NH₃ Cleanup	[illegible]		[illegible]	
Sulfur Cleanup	[illegible]	[illegible]	[illegible]	[illegible] (0.57%)
Steam Cycle Plant Auxiliaries	[illegible]	[illegible]	[illegible]	[illegible],819 (0.88%)
Gas Turbine Plant Auxiliaries	[illegible]	[illegible]	[illegible]	[illegible] (0.40%)
Net Plant Output - Mw	[illegible]	[illegible]	[illegible]	1,071.9

*Indicates Latent Heat From Condensation of Gasifier Steam the gas stream on overall efficiency. The actual heat lost as compared to a high temperature system with no gas cooling is the indicated value less the contribution due to condensation of the gasifier steam (given in parenthesis). A similar accounting is followed in the cleanup process. Because the gasifier was charged with the energy required to raise the inlet air from 59 F to gasifier inlet temperature, the heat of compression of the gasifier air must be credited to the system. This is done in the same manner using the ratio of steam to combined-cycle efficiency. The result is an effective fuel energy which, when used at combined-cycle efficiency, will produce the gross plant output.

While this approach may penalize a system more heavily than is warranted for the use of low-grade steam, it does provide a guide to determining the areas requiring attention. After subtracting compressor power and accounting for generator efficiency, the total electrical output corresponds to total generator output (computer calculations) within 1 percent. The auxiliary power requirements are separated to show the amount consumed by each part of the system.

The effect of this approach to bookkeeping is to reduce the penalty associated with the use of heat extracted from the steam cycle as opposed to either chemical or sensible energy from the fuel gas stream. Therefore, whereas the BCR/Selexol and BCR/Consol Systems have almost the same gasifier net thermal requirements (see net fuel energy leaving gasifier in Table 43, which is a straight thermal energy accounting) the effective fuel energy output at combined-cycle efficiency is significantly higher for the Consol System. This occurs because there is no need to cool the fuel leaving the gasifier and the fuel sensible energy is used at combined cycle efficiency. Although the resultant net heat required by the gasifier is quite high, the performance penalty associated with the use of thermal energy from the steam cycle is low in comparison to the sensible heat in the fuel gas.

The advantages of high temperature in the cleanup can be readily evaluated by comparing the BCR/Selexol and BCR/Consol Systems. The benefit of high temperature shows up in the gasifier performance. With the same coal input rate, the effective gasifier output is some 400 MM Btu/hr greater in the high-temperature system than in the low-temperature system. This 400 MMBtu/hr corresponds to a 7 percent increase in net output power and compares to an overall 18 percent increase in net output power for the high-temperature system. The remaining difference is due to a more efficient sulfur removal system and the lack of the process requirements of an ammonia scrub. The difference in sulfur removal systems accounts for 7 percent and the ammonia removal for another 4 percent of the overall difference in net output. Thus, while the high fuel temperature is important, it is not the major contributor to the improved

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performance of the BCR/Consol System. In fact, if the low-temperature cleanup system were as efficient. (i.e.. had similar process requirements) performance of the two could be made to match closely by regeneration and resaturation with water vapor of the fuel gas stream.

While a performance value for sulfur removal has not been set, the Selexol System offers the advantage of 95 percent + sulfur removal as opposed to a concentration of 540 ppm in the Consol product gas. The reason for the high auxiliary power and process heat in the Selexol process is to permit virtually complete removal of COS. If it were designed only for H_2S removal, the sulfur level in the product gas would rise to about 700 ppm and the utilities would drop to the level of the Bu Mines/Selexol System. This is within about 1 percent of the Consol System in terms of net output power.

Comparing the two gasifiers with low-temperature cleanup, the advantage of BCR-type over the Bu Mines-type can be attributed to the difference in sensible heat rejected, amounting to 262 MM Btu/hr, and the latent heat in the gasifier steam (75 MM Btu/hr) utilized in the BCR system. When combined, these total 4.3 percent of the fuel input, the exact difference in effective gasifier efficiency. The reasons for the difference in Selexol system performance has already been discussed. The major factor in the higher performance of the BCR/Selexol over the Bu Mines/Selexol System is, of course, the turbine inlet temperature rather than gasifier or cleanup performance.

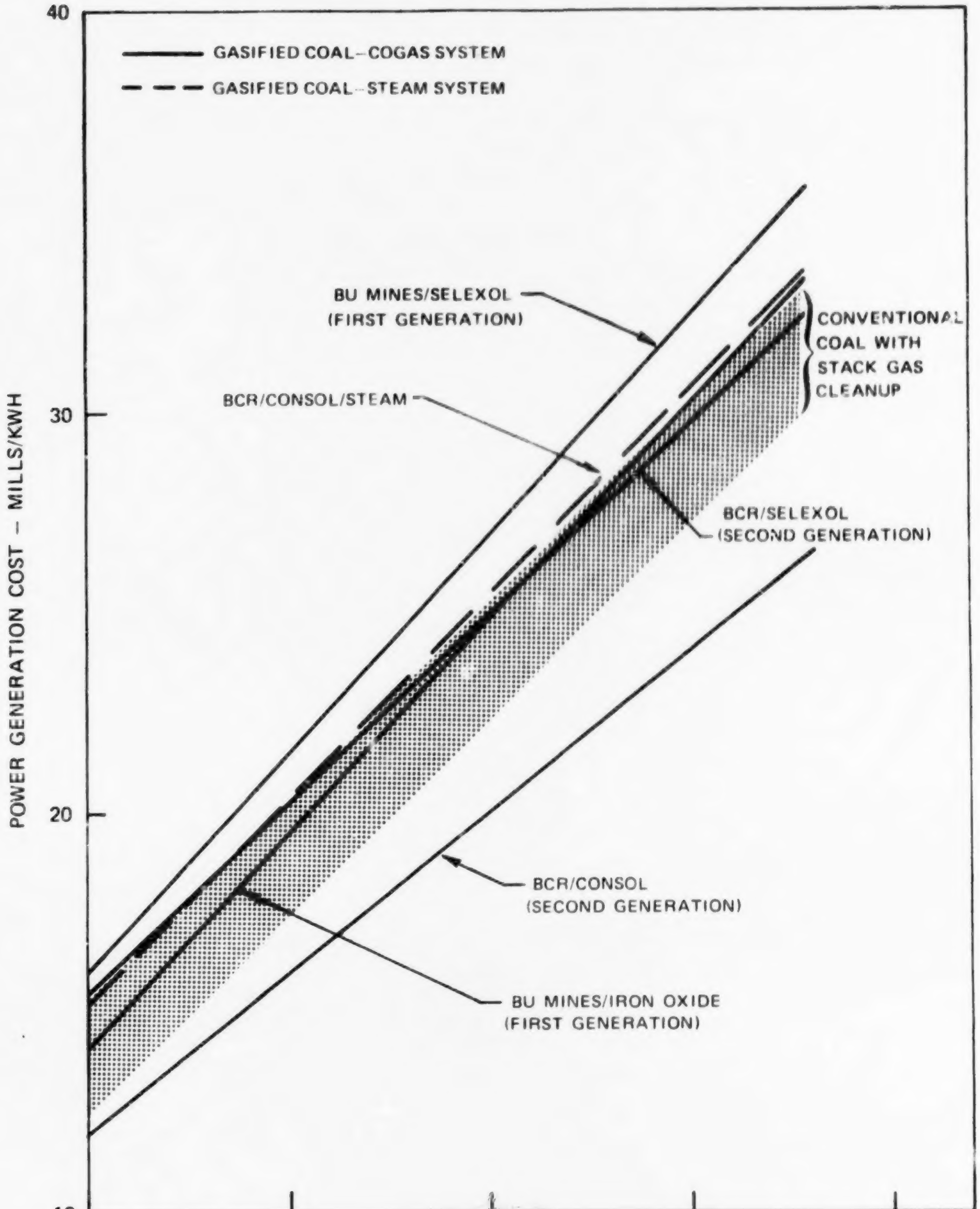
Considering the two Bu Mines-type systems, the high-temperature cleanup has a considerable effect on gasifier performance raising the effective efficiency by some seven (7) points. However, overall system performance shows little improvement over the low-temperature systems due to the inefficient sulfur recovery system. By producing sulfuric acid as opposed to elemental sulfur as in the other systems, significant improvement might result since the high utilities are entirely due to the production of SO_2 in the iron oxide regeneration process.

COST SUMMARY

For each of the four selected systems, an estimate of capital and operating cost has been made. In addition, the cost of power generation in a conventional plant with stack gas cleanup and in a gasified coal-fired steam plant has been estimated to provide a basis for comparison. A summary of total power generation cost for the six systems is presented in Figure 64. These are based on mid-1974 costs with escalation through the construction period. It is quite clear that the most attractive gasification/cleanup system for use with a COGAS powerplant is the BCR/Consol combination. This

FIG. 64

COST SUMMARY



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second-generation combination appears capable of generating electric power at much lower cost than any other alternative, including steam with gasification and steam with stack gas cleanup. The other gasification/COGAS system combinations are only marginally competitive with the steam alternatives. Unlike a nuclear system where fuel cost is relatively insignificant, capital cost effects will not be negated by changes in coal cost and the relative cost comparisons do not change greatly over the coal cost range of interest. Considering the zero coal cost intercept, which is a rough indication of capital cost, it is seen that a 30 percent increase in cost results in changing the BCR/Consol power system from COGAS to steam.

Also, for a given gasifier and power system, the low-temperature cleanup is more costly than the high. This is mostly due to the lack of ammonia removal and associated equipment in the high-temperature cleanup systems. However, in the case of the BCR/Consol system that effect is amplified by the improved efficiency of the gasifier-cleanup combination which results in a greater plant output over which to amortize the capital cost. Changing from first- to second-generation systems also tends to decrease initial cost due to a higher gas turbine specific output resulting in a lower per kilowatt cost of the power system.

The proportion of total cost ascribed to the various major system elements are given in Tables 48-50. These assume an owning cost of 17 percent of the capital investment spread over a load factor of 70 percent. On a straight capital cost basis, the coal-fired plant with stack gas cleanup is virtually as expensive as any of the gasified coal - COGAS systems. However, the estimated operating costs are about [illegible] mils more for the low-temperature systems as opposed to the conventional steam system. The absolute magnitude of this difference is subject to a fair degree of uncertainty. Lacking specific data on operating costs for the gasifier and cleanup system, they were taken to be 8.5 percent of the capital cost. For steam plants this factor runs about 2 percent and for combined-cycle plants about 3.5 percent. The 8.5 percent was chosen to be somewhat conservative since process plants generally run slightly lower than this.

A major factor that must be remembered in comparing costs is the difference in the form of the sulfur by-product, the variation in sulfur removal efficiency, and NO_x production which is certainly better in low-temperature systems where all ammonia is scrubbed from the product gas.

All of the gasified coal plants include provision for recovery of elemental sulfur. No credit has been taken for recovery of this resource, nor has credit (debit) been given to the disposal of the by-products of stack gas cleanup.

Table 48

INTEGRATED SYSTEM COST SUMMARY

	First Generation		Second Generation	
	BuMines Selexol	BuMines Iron Oxide	BCR Selexol	BCR Consol
Capital Costs - $/kw				
Power System Cost	232	230	208	190
Gasification System Cost	111	107	117	99
Cleanup System Cost	88	48	89	35
Total Plant Cost	431	385	414	324
Owning-Plus-Operating Costs-mills/kwhr				
Owning Costs (17% of Capital)	11.94	10.66	11.47	8.97
Operation & Maintenance				
Power System	1.32	1.31	1.19	1.08
Gasification & Cleanup	2.75	2.14	2.84	1.85
Fuel Cost at 60¢/MM Btu	6.52	6.40	5.69	4.82
Total Cost of Power	22.53	20.51	21.19	16.72

Table 49

COST SUMMARY FOR
COAL-FIRED STEAM WITH STACK GAS CLEANUP

Capital Cost - $/kw	
Coal Fired Plant	345
Stack Gas Cleanup	81
Total Capital Cost	426
Owning-Plus-Operating Costs-mills/kwhr	
Owning Cost (17% of Capital)	11.8
Operation & Maintenance	
Steam System	1.1
Stack Gas Cleanup	1.1
Fuel Cost at 60¢/MMBtu	5.8
Total Cost of Power	19.8
Net Plant Heat Rate - Btu/kwh	9,721

Table 50

COST SUMMARY FOR
BCR/CONSOL STEAM-CYCLE POWER PLANT

Capital Cost - $/kw	
Steam Plant	252
Let-Down Turbine	28
Gasifier	111
Cleanup	44
Total Capital Cost	435
Owning-Plus-Operating Costs-mills/kwhr	
Owning Cost (17% of Capital)	12.0
Operating Cost	
Steam Plant	.8
Gasification & Cleanup	2.5
Fuel Cost at 60¢/MMBtu	6.1
Total Power Cost	21.4

In preparing the capital and operating cost estimates for these systems the degree of detail with which each part of the system was evaluated varies with the available data. As a result, the confidence level is not the same in the various parts of one system or between different systems. However, the various component costs are treated consistently and the resultant comparisons are believed to be valid.

For each system, direct construction costs for the individual sections were identified which include all equipment and installation costs. To this is added an [illegible] percent contingency factor and 15 percent for engineering and supervision. Escalation is applied at 7.5 percent assuming a linear rate of expenditure. The same approach is used for interest during construction which is taken to be 10 percent per year. Using the "General Cost Model" (50) as discussed in Appendix C, the costs can be related to specific geographic regions. Unless otherwise noted, all cost summaries are typical of an Illinois, Ohio, Pennsylvania region.

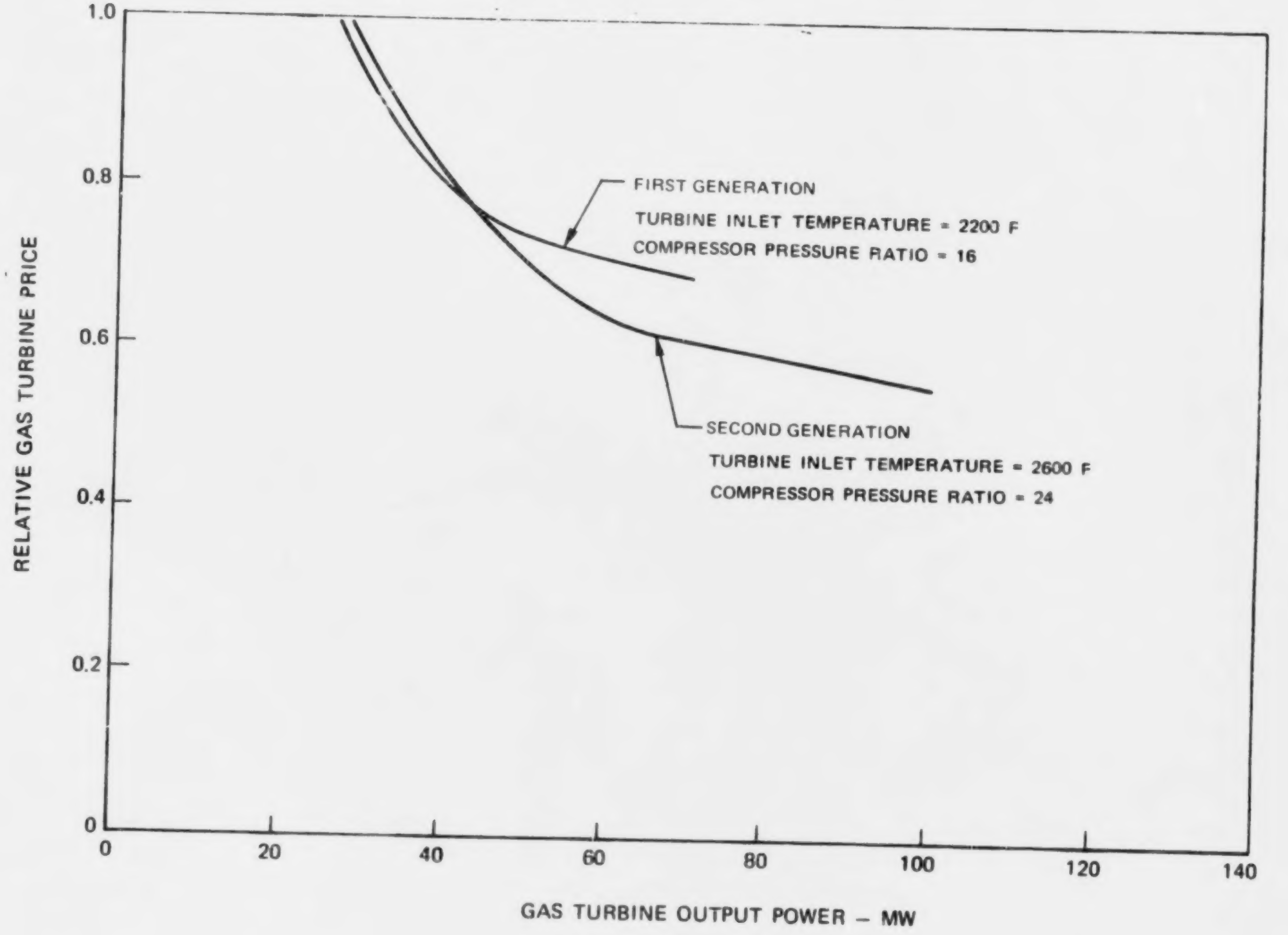
Gasified Coal - [illegible] System

The costing procedures described in detail in Appendicies C and I were followed in estimating gas turbine manufacturing cost and in determining powerplant costs. The gasifier and cleanup sections were treated like subsystems of the powerplant. While it is difficult to assign responsibility for the cost of the boost compressor and heat exchange equipment located in the gasifier air supply stream, these costs have been charged to the gasifier in the summary charts but have been carried as a separate item in the cost tabulations. Gasifier and cleanup system utility requirements are treated as an addition to the pertinent part of the main power system. The cost of feedwater treatment, cooling tower and condensate polishing equipment are calculated on the basis of their marginal or incremental cost which is then assigned to the appropriate subsystem. Since there is no provision for waste water treatment in the main power system an allowance of 4 percent of total gasifier and cleanup system cost was made for that function. This cost was assigned to the cleanup system.

Because plant size was determined by the net output resulting from a fixed ([illegible],000 lb/hr) coal firing rate, equipment size considerations would generally result in some modification of that firing rate. However, this modification would be slight since plant size is generally sufficient to accommodate [illegible] gas turbine engines designed for maximum output power corresponding to their turbine inlet temperature. Two steam turbines would be used when steam turbine size is greater than 100 Mw. Figure [illegible]5 shows the variation in gas turbine per kilowatt costs as a function of size and operating conditions. For each generation the minimum cost engine was used, that being 70 Mw for

FIG. 65

EFFECT OF OUTPUT POWER ON GAS TURBINE PRICE



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first generation systems and 100 MW for second generation systems. Engine costs for these two designs were determined using the procedure described in Appendix D.

The detailed compilation of power system costs are presented in Table 51 using the Bu Mines/Selexol system as an example. Since the powerplant cost data are for the East Coast (New York, New Jersey) area, the General Cost Model discussed in Appendix C was used to transpose the resultant cost to the Central (Illinois, Ohio) area. For the Bu Mines/Selexol System this is shown in Table 52 which gives the relationship between the two areas and also the total power section costs escalated through a four year construction period. Costs for the power system sized to operate with each of the four gasifier and cleanup combinations are summarized in Table 53.

Gasifier and cleanup system costs are summarized in Table 54. These were estimated using currently available cost data. For lack of cost data on the high-temperature desulfurization processes, investment costs were developed from conceptual plant designs for the Consol and Sintered Iron Oxide processes. Other sections were estimated from cost curves developed from published data.

Considering the sources from which the costs were developed, an accuracy of not better than ± 5 percent should be expected. Of the total investment costs, the coal gasification system represents about half. The available data on fixed-bed gasification systems costs are very sparse since no commerical plants have yet been constructed in the U.S. Reliable data for second-generation gasifiers are essentially non-existent since these systems are only in developmental stages where preliminary estimates have been made on conceptual pilot plant designs.

The only significant conclusion that can be drawn from the cost figures is that, for a given gasification system, the low-temperature cleanup scheme is more expensive, possibly by 20 percent. The costs for the desulfurization sections appear to be comparable for both the low- and high-temperature schemes. The former scheme must also bear the additional costs associated with gas cooling, sour water stripping, and ammonia recovery.

Coal-Fired Steam Station

A cost estimate was prepared for the reference coal-fired plant described in Section 1. Capital costs are summarized in Table 55. These are based on previous work (3) and have been escalated to a mid-1974 time period and appropriate interest and escalation added for a four year construction period. Stack gas cleanup system costs were added. For the mid-1974 time frame a

Table 51

POWER SYSTEM COST DETAILS

BuMines/Selexol/COGAS

Account 341	Site Preparation	$848,250
341-18	Administration Building	563,830
341-19	Turbogeneration Building	3,948,000
341-20	Tank Farm	1,236,100
341-23	Condensate Polishing System	800,640
341-24	Stack	362,000
	Total 341:	$7,758,820
Account 343		
343-01	Gas Turbine (8) Includes Installation, Labor	$17,643,500
343-02	Starter Motor (8)	85,000
343-03	Torque Converter	804,000
343-04	Lube Oil Purifier & Storage (Pumps, Filters, Etc.)	224,000
343-05	Lube Oil Fire Protection	160,000
343-07	Air Compressor Services, Instrumentation	140,000
343-08	Breeching	2,140,600
343-09	Expansion Joints; Not Applicable in COGAS Plant	0
343-10	Inlet Air Filters	555,760
343-12	Energy Cooling Tank Pump & Piping	11,200
343-13	Fuel Oil Heaters & Pumps	123,200
343-14	Miscellaneous Pumps & Tanks	56,000
343-15	Control Panels	560,000
343-16	Computer Controls	560,000
343-17	Fuel Piping	1,249,400
343-18	Fuel Pipe Insulation	187,500
	Total 343:	$24,500,760
Account 344	Generator For Gas Turbine	$10,537,630

Account 312

312-01	Waste Heat Boiler	$22,432,400
312-02	Boiler Feed Pump	339,390
312-03	Boiler Feed Tank Deaerator	133,070
312-04	Water Treatment (Demineralization)	691,960
312-05	Condensate Storage Tank	29,940
312-08	Miscellaneous Pumps	70,525
312-09	Piping	3,080,240
312-10	Insulation for Piping	264,420
312-11 312-12	Controls Computer } Steam Turbine Only	308,025
	Total 312:	$27,349,970

Account 314

314-01	Steam Turbine and Generator (Output per Unit - 105,800 kw)	$ 9,833,270
314-03	Condenser & Tubes	1,212,180
314-04	Condensate Vacuum Pump & Motor	134,795
314-05	Condensate Pump & Motor	274,780
314-06	Cooling Tower	5,994,080
314-08 314-09 314-10	Circulation water Valves & Expansion Joints Circulation Water Pumps Make-Up Structure: Screens & Pumps	1,146,280
	Total 314:	$18,595,385

Accounts 345 & 353

Accessory Electrical Equipment	$8,768,750

Account 346

Miscellaneous Power Plant Equipment	$359,040
Other Expenses	$1,957,410
Total Direct Construction Costs	$99,827,765
Contingency (8%)	$7,986,220
Engineering & Supervision (15%)	$14,974,165
Total Unescalated Cost	$122,788,150

Table 52

EFFECT OF LOCATION ON POWER SYSTEM COST
BuMines - Selexol - COGAS

	East Coast	Central
Materials & Equipment	$61,244,800	$61,244,800
Wages, Fringes, Administration & Supervision	26,355,680	22,954,040
Engineering	9,186,720	9,186,720
Tools During Construction	2,878,505	2,878,505
Total Direct Construction Costs	99,665,705	96,264,065
Contingency	9,966,570	9,626,405
Profit & Insurance	13,155,875	12,706,855
Total Before Escalation	$122,788,150	$118,597,325
Escalation	20,595,815	19,892,870
Interest During Construction	33,272,250	32,136,650
Total Capital Cost	$176,656,215	$170,626,845

Table 53

COGAS POWER SYSTEM COST SUMMARY

FPC Account Number		BuMines/ Selexol	BuMines/ Iron Oxide	BCR/ Selexol	BCR/ Consol
		Costs - $1,000			
341 -	Structures and Improvements	7,494	7,494	11,011	11,422
343 -	Prime Movers (Gas Turbine)	23,665	24,637	20,725	23,556
344 -	Electric Generators (Gas Turbine)	10,178	10,315	9,514	10,077
312 -	Boiler Plant Equipment	26,417	26,204	29,445	32,360
314 -	Steam Turbine Generator Units	17,961	18,148	23,230	22,822
345 & 353 -	Accessory Electrical Equipment	8,469	8,679	10,350	12,009
346 -	Miscellaneous Power Plant Equipment	347	351	385	416
	Other Expenses	1,891	1,917	2,093	2,253
	Direct Construction Costs	96,421	97,749	106,754	114,914
	Contingency, Engineering & Supervision	22,177	22,482	24,553	26,430
	Total Construction Costs	118,597	120,231	131,308	141,344
	Interest & Escalation	52,030	52,746	57,606	62,009
	Total Capital Cost (Power System Only)	170,627	172,978	188,913	203,353

Table 54

GASIFIER & CLEANUP SYSTEM CAPITAL COST BREAKDOWN

	BuMines/ Selexol	BuMines/ Iron Oxide	BCR/ Selexol	BCR/ Consol
Gasification	62.42	62.43	83.23	83.23
Gas Cooling	14.86	-----	23.78	-----
Desulfurization	23.78	20.81	29.72	20.81
Sour Water Stripping	5.94	-----	5.94	1.49
Ammonia Recovery	10.4	-----	8.92	-----
Sulfur Recovery	2.97	10.4	2.97	8.92
Waste Water Treatment	4.82	3.75	6.18	4.58
Boost Compressor & Boiler	10.8	10.72	11.0	11.73
Feedwater Treatment	6.73	5.8	9.42	9.42
Cooling Tower	1.06	-----	1.75	.42
Condensate Polishing	.06	.22	.25	.03
Other Expenses	2.88	2.28	3.66	2.81
Total Capital Cost (Includes Escalation & Interest)	146.74	116.41	186.83	143.44

Table 55

COST SUMMARY - COAL-FIRED STEAM PLANT

1000 Mw Nominal Size

FPC Category	Cost - $1,000
311 - Structures & Improvements	21,604
312 - Boiler Plant Equipment	91,752
314 - Turbogenerator Units	53,906
315 - Accessory Electric Equipment	16,032
316 - Miscellaneous Power Plant Equipment	792
353 - Station Equipment	2,329
Other Expenses	3,728
Direct Construction Costs	190,143
Indirect Costs	43,733
Total Construction Cost	233,876
Escalation and Interest	96,444
Total Capital Investment	330,320

total investment cost of $81/kw was used.(51) Another reference,(52) gives capital costs ranging from $58/kw for a limestone system to $90/kw for a Wellman-Lord system. However, it is generally quite difficult to determine what corrections must be made to these numbers to put them on an equal basis with those in this report. Reference 32 places the cost of a magnesia slurry regenerable system at about $44/kw for a 1972 time period. When adjusted to 1974 and using an approach to interest and escalation that is consistent with this study, this increases to about $68.4/kw.

The overall effect of stack gas cleanup on power cost is about 4.16 mills per kwh. This is made up of some 2.7 mills capital costs, 1.2 mills operating costs and 0.3 mills additional fuel costs. Because of the wide variations that are reported in the literature, the coal-fired unit costs have been presented as a band in Figure 64, with regenerable systems represented by the high side and nonregenerable limestones systems falling on the low side of the band. As an overall check, the data in Ref. 32 was used to calculate the cost of a magnesia slurry system under the ground rules selected for this study. The resultant cost was 3.9 mills per kwh , as compared to 4.16 for this study. Reference 52 gives overall costs of 3.68 and 4.66 mills per kwh for nonregenerable and regenerable systems respectively.

Gasified Coal-Fired Steam Plant

The BCR/Consol/Steam System previously discussed was used to evaluate the cost aspects of this type of system. This combination represents the highest gasifier/cleanup system performance and lowest capital cost per unit output. Gasifier and cleanup system costs have been presented earlier in this section. The low Btu gas-fired steam system costs(3) were escalated as required. The other major parts of this system are the air compressor, let-down turbine, and the boilers and intercoolers located in that stream. Because of the relatively high pressure ratio for the let-down turbine-compressor combination, it was treated as two low-pressure ratio engines in preparing the estimate. A breakdown of the total cost is presented in Table 56.

Table 56

BCR/CONSOL/STEAM PLANT COST SUMMARY

	Cost $1,000
Gasifier & Cleanup System (Less Boost Compressor & Boiler)	131,730
Low Btu Steam Plant	214,187
Compressor-Let Down Turbine (Including Heat Recovery)	23,795
Total Capital Investment (Includes Interest & Escalation)	369,712
Plant Output Power - Mw-	850
Unit Cost - $/kw	435

SECTION 5

ENVIRONMENTAL IMPACT AND GOALS

The environmental impact of the integrated power system and the cleanup system goals required to meet environmental and operational constraints are identified in Section 5. Also, included in this section are discussions of effects of low- and high-temperature cleanup processes on the economics and performance of the selected power systems, the technology needed to attain the selected goals, and further research and development required to meet these goals.

IDENTIFICATION OF ENVIRONMENTAL IMPACT

The four integrated power systems defined in Section 4 were used as a basis to identify the various pollutants which would be associated with the gasification system and the power system. A summary of the estimated environmental impact of these four systems is given in Table 57. In the following paragraphs, a more detailed description is given for each of the pollutants quantified in Table 57, as well as a qualitative description of other pollutants which would be associated with the various processes.

Air Emissions from Gasifier/Cleanup Systems

It is apparent from Table 57 that the SO_2 emissions from the sulfur recovery plant could be significant (60 percent to 85 percent of the total integrated sulfur emission). However, commercially available tailgas treating processes may be used to further reduce sulfur levels in these cases. For example, the addition of an alkali scrubbing system to treat the sulfur plant tailgas can achieve >99 percent overall sulfur recovery. Tailgas treatment in the BCR/Consol case will not significantly reduce the sulfur emission since 94 percent of the SO_2 results from combustion of the fuel gas product.

Effluent streams from the battery limits plant producing clean, low-Btu fuel gas from coal are summarized in Table 58 for the four integrated gasifier-gas purification combinations selected for standard system analysis.

Table 57

ESTIMATED ENVIRONMENTAL IMPACT OF INTEGRATED POWER SYSTEMS
(All Values in lb/10^6 Btu Coal Unless Otherwise Noted)

	Waste Water	Thermal(1)	SO_2	NO_x(10)	Particulate	Solids	Sulfur(2)	Ammonia(2)
BCM-Selexol	219(3)	0.005	0.320(6)	0.356	-----	14.6	2.96	1.38
2200 F COGAS	27(3)	-----(5)	0.088	0.174(7)	0.01(8)	----	----	----
Total	246	0.005	0.408	0.530	0.01	14.6	2.96	1.38
BOM-Iron Oxide	---	0.018	0.535(6)	0.31	----	14.6	2.51	----
2200 F COGAS	27(1)	----(5)	0.337	3.79(7)(11)	0.06(9)	----	----	----
Total	27	0.018	0.87	4.10	0.06	14.6	2.51	
BCR-Selexol	520(3)	0.016	0.487(6)	0.300	----	7.1	2.74	0.96
2600 F COGAS	24(4)	-----(5)	0.080	0.012(7)	0.01(8)	----	----	----
Total	544	0.016	0.567	0.312	0.01	7.1	2.74	0.96
BCR-Consol	75(3)	0.002	0.033	-----	-----	7.1	2.72	-----
2600 F COGAS	22(4)	-----(5)	0.520	2.65(7)	0.04(9)	----	----	
Total	97	0.002	0.553	2.65	0.04	7.1	2.72	-----

(1) Millions of Btu's rejected per million Btu's of coal
(2) Recovered with no credit for subsequent sale
(3) See Table 60 for water characteristics
(4) Includes boiler make up and cooling tower losses
(5) Mechanical draft cooling towers are used
(6) Includes emissions from Claus plant. A 90%+ reduction could be obtained with tail-gas cleanup
(7) Fuel-related NO_x only
(8) Assumed to be equal to methane-fired systems
(9) Sum of carryover and combustion products
(10) All weights as NO_2
(11) Includes nitrogen in tar

Table 58

GASIFIER/CLEANUP SYSTEM EFFLUENT SUMMARY

Gasifier System	Bureau of Mines		BCR	
Cleanup System	Selexol	Iron Oxide	Selexol	Consol
Elemental Sulfur, LTPD	259.1	219.8	258.1	256.7
Anhydrous Ammonia, STPD	133.2	-----	99.2	-----
Ash/Slag, STPD	1369.6	1369.6	730.8	730.8
Other Solids, STPD	163.4	249.6	-----	139.8
Waste Water, GPM	351	-----	477	129
Stack Gas (Sulfur Recovery Plant)				
mols/Hour	7728	31,191	13,051	1,613
M.W.	29.85	28.95	32.67	28.18
Vol.% N_2	59.60	75.77	49.37	97.03
O_2	2.77	7.78	2.15	2.34
CO_2	19.80	8.25	35.12	-----
SO_2	0.52	0.21	0.50	0.28
NO	0.80	0.17	0.43	-----
H_2O	16.51	7.82	12.43	0.35

In addition to the SO_2 emissions from the stack gas, there will be NO_x. The NO_x is due to both thermal NO_x from the incineration process and also to the carryover of fuel nitrogen compounds, such as NH_3, in the sulfur recovery plant feed gas. Estimates for the SO_2 with and without stack treatment and NO_x emissions are given in Table 59.

The only particulate emission which can be associated with the gasifier/cleanup systems would be those due to the disposal of ash/slag or spent solid absorbent. No estimates have been made for any entrainment of these products during disposal.

Water Emissions From Gasifier/Cleanup Systems

The aqueous waste streams from the low-temperature cleanup systems are potentially hazardous to the environment. These streams will contain non-strippable contaminants originally present in the raw producer gas. Data on the nature and quantity of trace contaminants in coal derived producer gas are not available. At best, a partial qualitative identification can be made based on information given in the literature.(53,54,55) Analysis of waste water streams given for specific coal gasification processes and coke plants are given in Table 60.

In addition, trace elements present in raw coal may be volatilized during the gasification process and subsequently scrubbed out in the water washing steps. An indication of the elements likely to be found in the water stream is given by the analysis of Illinois coals (56) and the process condensate from gasification of an Illinois No. 6 coal via the Synthane process (53) as presented in Table 61. Of particular concern are those elements named as hazardous to human health by the EPA (57): beryllium, fluorine, arsenic, selenium, cadmium, mercury, and lead. These elements are all volatile and can be expected to appear in the raw producer gas and ultimately in the waste water stream. These elements may pass through high-temperature cleanup systems essentially untouched and thus appear in the stack gas much the same as from a conventional coal-fired steam plant.

The high operating temperature of the BCR gasifier, chosen for second generation use, would preclude the appearance of phenols, tars, and fatty acids in the raw gas and consequently in the waste water. Aqueous wastes, from the first generation system should contain these compounds and traces of heavy metal contaminants. After steam stripping, the waste water from this type system could conceivably contain the constituents given in Table 62.

Design of a water treatment system to render the aqueous wastes acceptable for discharge to the environment would require detailed analysis for the specific gasification system involved. Such design is beyond the scope of the current study.

Table 59

AIR EMISSIONS FROM GASIFICATION/CLEANUP SYSTEMS

	SO_2 lb/MMBtu	NO_x lb/MMBtu
[illegible]/Selexol		
Stack Gas W/O Cleanup	0.540	.326
Stack Gas/Cleanup	0.032	----
[illegible]/Iron Oxide		
Stack Gas W/O Cleanup	0.535	0.310
Stack Gas/Cleanup	0.038	----
BCR/Selexol		
Stack Gas W/O Cleanup	0.487	0.300
Stack Gas/Cleanup	.036	----
BCR/Conosol		
Stack Gas W/O Cleanup	0.033	----

Table 60

WASTE-WATER STREAM ANALYSIS
(ppm)

Constituent	Coke Plant	Coal Gasification Plants	
		Synthane	El Paso /Lurgi
Total Ammonia	1,800-4,300	7,000-10,000	1,050
Total Sulfur	0-50	1,400	
Phenol	410-2,400	2,600-6,600	500
Thiocyanate	100-1,500	20-200	
Cyanide	10-37	0.1-0.6	
Fatty Acids			1,750
Chlorides		500	
Carbonate	1,200-2,700	17,000	
COD	7,500-10,000	15,000-38,000	
pH	8.3-9.1	8.6-9.2	

Table 61

TRACE ELEMENT ANALYSIS OF ILLINOIS COAL

Element	Illinois Coal*	Waste Water**
Al	1.29 %	1,000 ppb
Ca	0.77	4,000
Cl	0.14	
Fe	1.92	3,000
K	0.16	160
Mg	0.05	2,000
Na	0.05	
Si	2.49	
Ti	0.07	
As	14.0 ppm	30 ppb
B	102.0	
Be	1.6	130
Br	15.4	
Cd	2.5	6
Co	9.6	2
Cr	13.8	6
Cu	15.2	20
F	60.9	
Ga	3.1	
Ge	6.6	30
Hg	0.2	
Mn	49.4	40
Mo	7.5	
Ni	21.1	30
P	71.1	90
Pb	34.8	
Sb	1.3	
Se	2.1	360
Sn	4.8	20
V	32.7	3
Zn	272.3	60
Zr	72.5	

* Mean value for 101 coals analyzed including some 19 non-Illinois coals. For Illinois coals only, EPA 650/2-74-054, July, 1974 contains a similar table for 82 different coals.

** Process condensate from gasification of Illinois No. 6 coal.

Table 62

POTENTIAL WASTE-WATER STREAM ANALYSIS FOR BOM GASIFIER

	ppm
Ammonia (free & fixed)	200-400
Carbonates	250
Sulfides	10-100
Phenols	500
Fatty Acids	1,800
Thiocyanates	100
Cyanides	1-10
Chorides	500
Heavy Metals	10-20
BOD	2,500
pH	9

Solid Wastes From Gasifier/Cleanup Systems

Both first- and second-generation systems will discharge substantial quantities of solid material for disposal to the environment, primarily ash and slag. With the coals selected, the BOM gasifier would produce about 1400 TPD ash and the BCR gasifier approximately 700 TPD of slag. First-generation, stirred-bed gasification systems will also discharge 160 TPD of ash and coal dust recovered from the raw producer gas via particulate removal systems. There is the possibility of briquetting a portion of this dust for recycle to the gasifier, providing the carbon content justifies the additional expense. In addition, there are chemical solid wastes generated in the high-temperature cleanup systems; 86 TPD of sodium sulfate from the sulfur recovery section in the first-generation iron oxide system, and 140 TPD of spent dolomite from the second-generation Consol system. The latter may be recycled if the activity for sulfur absorption can be restored after cold regeneration.

By-product sulfur from all systems, 200-250 LTPD, should present no significant environmental problems. In the elemental form, it can be stockpiled with little or no environmental hazard and is readily marketable. Similarly, the ammonia by-product, 100-130 STPD, from low-temperature cleanup systems can be stored in the liquid form with no associated pollution hazard and sold for its chemical value. In the present study, no credit has been taken for by-product sales.

Air Emissions From Power Systems

The cleanup systems used for the four systems were easily able to remove the H_2S in the low-Btu gas to levels which are a fraction of the EPA standard for coal-fired power stations (1.2 lb SO_2/MMBtu in coal), the low-temperature systems could potentially achieve removal to levels that are somewhat lower than the natural gas standard (0.1 lb SO_2/MMBtu). It was assumed that the H_2S remaining in the fuel gas was converted to SO_2 during the combustion process. Estimates for the SO_2 emissions are given in Table 63 and Figure 66 shows SO_2 emissions as a function of H_2S in 150 Btu/ft^3 gas.

Also shown in Table 63 are the NO_x emissions. As was discussed in Section 2, the total amount of NO_x emitted is a function of combustion temperature and time, the "thermal" NO_x, and of the nitrogen-bound fuel constituents. Thus, the low-temperature cleanup systems with their associated water washes and solvents which also remove NH_3; (the main fuel nitrogen compound produced during coal gasification) have low NO_x emissions in which the thermal NO_x could play a significant role. Those systems using high-temperature sulfur removal, however, let the fuel nitrogen compounds pass through, giving rise to NO_x emissions beyond the acceptable EPA limit for coal-fired stations (0.7 lb NO_x/MMBtu

Table 63

POWER SYSTEM EMISSIONS
(lb/MMBtu)

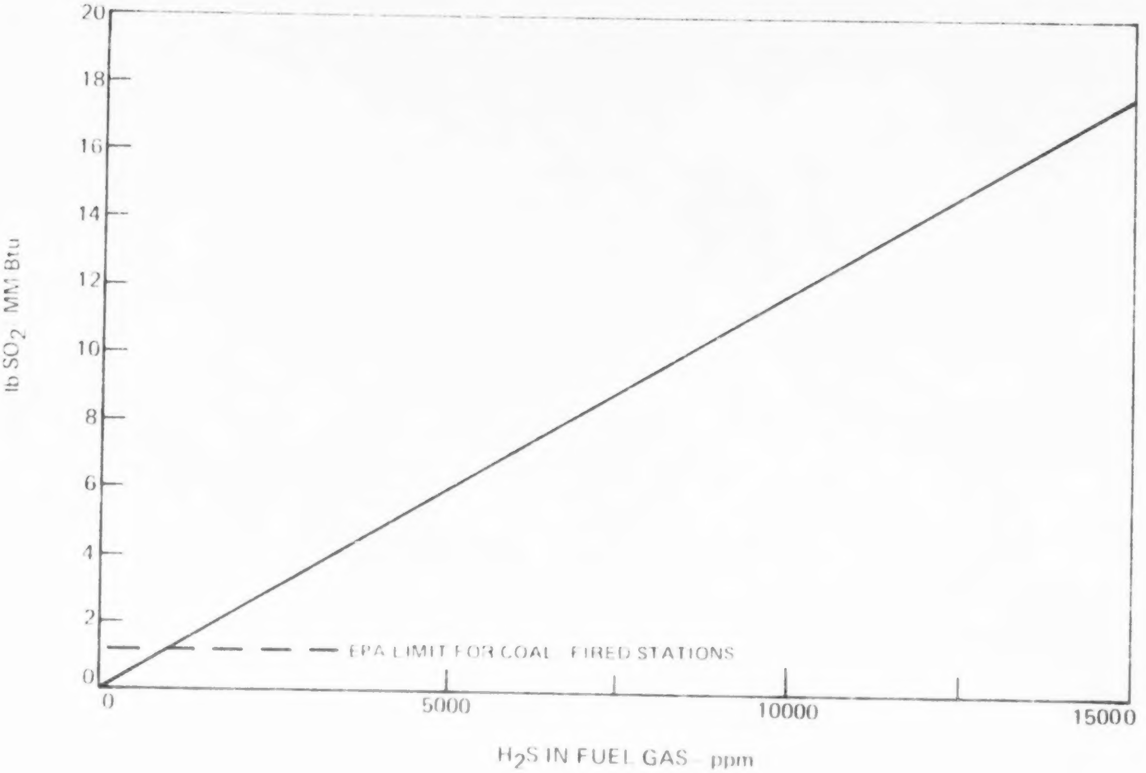
	SO_2	NO_x	Particulates
BOM/Selexol	0.058	0.174	<0.01
BOM/Iron oxide	.337	3.79	<0.06
BCR/Selexol	0.060	0.012	<0.01
BCR/Consol	0.520	2.65	<0.04

Thermal NO_x is not included in tabulation

FIG. 66

SO_2 EMISSIONS AS A FUNCTION OF H_2S IN FUEL GAS

HHV = 150 Btu/ft^3



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coal). In order to achieve environmental acceptability with the higher performance systems using the high-temperature sulfur removal systems, it will be necessary to identify a method of also removing fuel nitrogen compounds at high temperatures.

The emission of particulates from these power systems is difficult to quantify. The very limited information (58) available on particulates in gasifier off-gas which has been scrubbed by a sulfur removal system indicates that the loadings vary from 0.25 milligrams/m^3 (1.55×10^{-8} lb/ft^3) in a K-T system, to 1.0 milligram/m^3 (6.2×10^{-8} lb/ft^3) in a pressurized Lurgi system, to 2.5 milligram/m^3 (15.5×10^{-8} lb/ft^3) in a Winkler system. Assuming that these loadings are typical and would pass through the combustion process without alteration, the particulate emissions would range from about 7.5×10^{-5} lb/MMBtu based on the low value (1.55×10^{-8} lb/ft^3 for BOM/Selexol) to about 10^{-3} lb/MMBtu based upon the highest value (15.5×10^{-8} lb/ft^3 for BCR/Selexol). These values are well below the EPA standard of 0.1 lb/MMBtu for coal-fired stations. In addition to the particulate carryover in the fuel gas, there are particulates formed during combustion. Measurements (59) on stationary turbines firing natural gas show emissions of the order of 0.001 g/ft^3 in the exhaust. This is equivalent to approximately 0.75×10^{-4} lb/MMBtu. Assuming the low-Btu fuel gas burns as cleanly as natural gas, the total emissions would range from 1.5×10^{-4} lb/MMBtu to slightly over 1×10^{-3} lb/MMBtu, again well below EPA standards.

For those systems using high-temperature cleanup, there will be a higher loading. The dust carryover for the BOM gasifier was given as 0.016 lb/lb coal (Table 2). This value, also assumed as carryover from the BCR gasifier, is equivalent to about 1.16 lb/MMBtu. To meet the EPA limit for coal-fired stations, 92 percent removal must be obtained. High-temperature cyclones have the potential for removing this mass of particles. It is not apparent that meeting the EPA particulate limit means that the fuel gas and its combustion products are clean enough for use in high-temperature turbines. This will be discussed in the section dealing with cleanup system goals. Discussions on advanced high-temperature removal systems such as panel bed filters indicate that this type of system could have the potential of 99 percent-plus removal efficiencies. This would be equivalent to approximately 0.01 lb/MMBtu. For purposes of this study, it was conservatively assumed that removal efficiencies of 95 percent could be attained for first-generation systems and 98 percent for second-generation systems giving rise to the values in Table 63.

Water Emissions From Power Systems

Because of the use of mechanical draft cooling towers, which eliminates thermal pollution, the only significant water effluent is from boiler and cooling tower blowdown. Boiler blowdown is a function of steam conditions and water purity and can range from 22 to 27 lb H_2O/MMBtu coal input. The type of pollutants associated with these streams has been discussed adequately elsewhere.(30)

DEFINITION OF CLEANUP SYSTEM GOALS

The goals to be attained by the cleanup systems selected in Section 3 of the study are based upon two constraints - the EPA promulgated emission standards for coal-fired powerplants and the operational limitations of the gas turbine. The latter is usually expressed as some type of specification for fuels for turbine use. In setting goals to be attained, the limitations proposed by system operating parameters must also be considered, allowing flexibility in both the goals and the operating conditions so that the effects of one on the other can be adjusted to minimize overall environmental impact.

EPA Standards

The EPA standards for coal-fired steam stations of 250 x 10^6 Btu/hr and above have been discussed in previous sections. The rationale behind using coal-fired steam station rather than gas turbine standards is twofold; first, the overall system is, indeed, coal-fired and the environmental impact is a function of the entire fuel processing, cleanup, and power systems; secondly, no firm gas turbine standards have been promulgated. It is instructional, however, to compare the coal-fired steam station, possible EPA gas turbine standards (60) and the gas turbine standards as viewed by the industry through the American National Standards Institute (ANSI) Committee for Gas Turbine Procurement.(61) These three goals are shown in Table 64. (NOTE: Both the EPA and ANSI emission standards are suggested and are subject to revision.)

From Table 64 it is seen that the gas turbine regulations as proposed by the EPA are essentially the same as those for steam stations. Those suggested by ANSI are considerably more lenient (except particulates) and are based upon the industry's recognition that the ability to burn "dirtier" (higher sulfur content) fuels must be developed if a viable utility market for gas turbines is to be maintained. Higher NO_x limits are supported because it is imperative that a dry combustor (no water injection) must be used since water consumption could become a very major concern in many areas. The current industry efforts in low-NO_x burner research indicate that the 120 ppmv value is attainable dry.

Gas Turbine Fuel Specifications

Gas turbine manufacturers have established stringent specifications for fuels to be burned in industrial applications. A summary of pertinent

Table 64

COMPARISON OF EMISSION STANDARDS FOR GAS TURBINE SYSTEMS
(lb/MMBtu)

Pollutant	Agency EPA(1)	EPA(2)	ANSI(3)
SO_2	1.2	0.3% S Weight - 1.2	.8% S Weight - 3.2
NO_x	0.7	55 ppm - 0.7	120 ppm - 1.53(4)
Particulates	0.1	10% Opacity	10% Opacity

(1) For coal-fired steam stations
(2) Suggested standards for gaseous fueled system subject to revisions
(3) Not endorsed by ANSI or the institutes or companies of its members
(4) Suggested 1980 standards would be 65 ppm or 1 lb/MMBtu

specifications is given in Table 65. The major problem in current utility use of gas turbines is hot end corrosion. This is a result of two factors, the presence of alkali metal salts in the fuel and of fuel sulfur. The corrosion agent is the alkali metal sulfate which attacks the oxidation resistant coatings. Once these coatings have been penetrated, rapid oxidation of the base alloy occurs. The result is that, although the turbine could withstand fuels having sulfur contents well above those allowed by environmental constraints, the presence of alkali metals reduces the allowable content to a much greater degree.

The presence of particulates also contributes to corrosion since any spalling of the coatings due to impingement could result in subsequent oxidation of the base alloy.

Particulate Loading - One of the major concerns in setting fuel specifications is the particulate content. As can be seen in Table 65, the particulate loadings are low, varying from less than 1 ppm to about 30 ppm. The reason for the low allowable loadings is the potential erosion of the turbine blades.

Although a number of attempts have been made to utilize coal directly in a turbine (62, 63, 64) all have failed to obtain reasonable machine lifetime due to excessive erosion. Yet, the actual limitations of particulate content or size distribution have not been adequately defined. In a previous UTRC study (65) on methods of cleaning emissions from jet engine test cells, it was shown that the mean diameter of the particles emitted from liquid-fuel burning engines was 0.1μ with over 99 percent-plus of the particles being less than 1μ. Since erosion is not a problem with these engines, it can be concluded that particles of 1μ or lower in the fuel or combustion products are not harmful to engines. In fact, it was pointed out in this study that the use of combustion additions to eliminate visible emissions resulted in agglomerated particles of greater than 1μ (above the size range which results in refraction). While deposition did occur, there was no erosion indicating that the soft compounds which left the airstream and impinged on the blades did not damage them.

Tests in the United Kingdom on fluid-bed combustion (66) also indicate that soft particles of quite large diameters (~50μ) do not appear to cause erosion. However, tests carried out in Australia (67) indicate the hard particles such as would result from coal combustion above the ash fusion point cause erosion at sizes above 6μ. Perhaps the most useful guide would be the work done in the petro-chemical industry on turbine erosion. (68) Results of this work are shown in Figure 67, where engine life is shown as a function of particle size and loading. The shape of Figure 67 indicates that the relationship between lifetimes, particulate loading and particulate size would be of the form

Table [illegible]

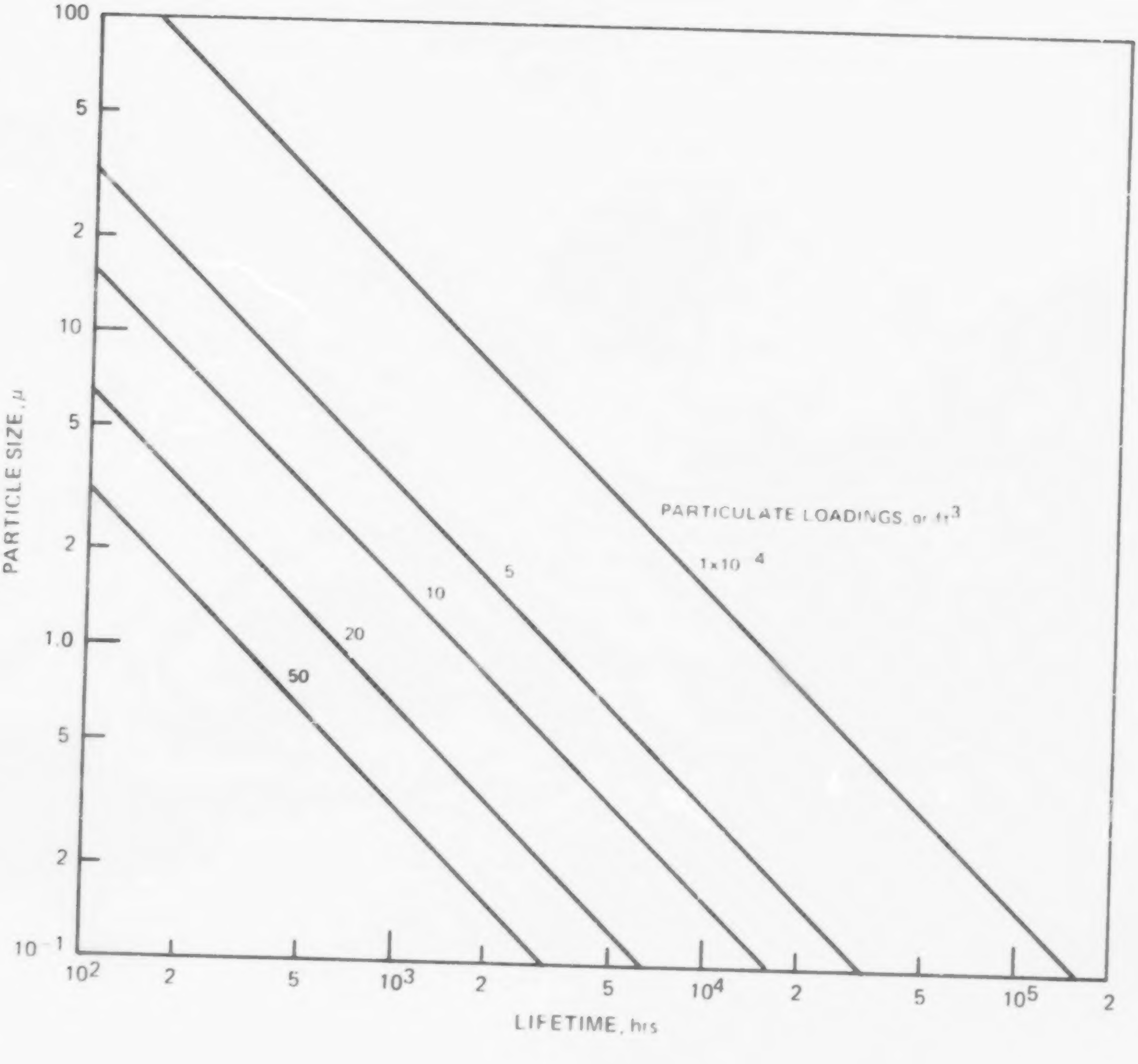
[illegible]

Constituent	[illegible]	Suggested(2)	[illegible](3)	General Electric(1)
Sulfur	[illegible]	[illegible]	[illegible] by Weight	Less Than Amount Required to form 5 ppm Alkali Metal Sulfates
Particulates	[illegible]	.01 gr/ft3 [illegible] Maximum	Limited by Material(4)	30 ppm (Weight) (0.1 g/ft3)
Metals				
Vanadium	[illegible] ppm (Weight)	-----	[illegible] ppm (Weight)	-----
Sodium and Potassium	[illegible] ppm (Weight)	See Sulfur Spec.	[illegible] ppm (Weight)	See Sulfur Spec.
Calcium	0.1 ppm (Weight)	-----	[illegible] ppm (Weight)	-----
Lead	0.1 ppm (Weight)	-----	[illegible] ppm (Weight)	-----
Copper	[illegible] ppm (Weight)	-----	-----	-----

(1) For aircraft-derived turbine using gaseous fuels
(2) For industrial turbines; subject to revision
(3) Liquid fuel specifications
(4) Given in Ref. 70 as 4 x 10^{-4} gr/ft^3 of 2μ to 10μ particulate in gaseous fuels

FIG. 67

EFFECT OF PARTICLE SIZE ON ENGINE LIFETIME



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$$L = \frac{\text{Constant}}{\mu \times C} \qquad (45)$$

where

L = lifetime

μ = particle size

C = loading

Thus, for a given lifetime, a reduction in particle size would be accompanied by higher allowable loadings, e.g., capture of particles of 2μ and above would allow twice the particulate concentration that capture of 4μ and larger particles would allow (assuming linear particle size distribution).

Suggested Low-Btu Fuel Gas Specifications

The values for loadings in Table 65 are based upon methane fuel. For low-Btu fuel, both the heating value and the density will affect the allowable loadings. For fuel gases of 150 Btu/ft^3, (about 2500 Btu/lb) the value in Table 65 would be reduced by approximately a factor of eight if the solids loading to the turbine were to be kept constant. Using the foregoing approach, Table 66 has been prepared to serve as a guideline for low-Btu fuel gas.

In addition to the usual fuel property limitations, Table 66 includes a value for fuel nitrogen compounds expressed in terms of NH_3. The limitation is based upon 90 percent conversion of fuel nitrogen to NO_x.

COMPARISON OF CLEANUP SYSTEM CHARACTERISTICS

One of the major considerations at the initiation of this study was the identification of the point(s) at which one type of cleanup process, either low- or high-temperature, demonstrated technical and economic benefit over the other. As the study progressed, it became apparent that when the power system, the gasification system, and the cleanup process were combined into an integrated power station, much could be done to reduce the differences between the various processes which would result from a cursory examination. The overall results of the power station integration have been fully documented in Section 4 of this report and the advantages/disadvantages of each system has been identified. However, the effect of the sulfur removal/recovery system alone on the performance and economics has not been defined.

To determine the penalties associated with the sulfur removal/recovery processes, each integrated system was broken into four parts; (1) a

Table 66

SUGGESTED LOW-BTU FUEL GAS CLEANUP SYSTEM GOALS

Property	Specification	Resulting Emission
Sulfur	0.05 Mol % or Less Than Amount to Form 0.6 ppm Alkali Metal Sulfates	0.44 lb SO_2/MMBtu
Particulates	4 ppm (weight) (0.0012 gr/ft^3)	<0.01 lb
Metals		
Vanadium	<0.03 ppm (weight)	
Sodium and Potassium	See Sulfur Spec	
Calcium	<0.012	
Lead	<0.012	
Copper	<0.0025	
Nitrogen Compounds	500 ppm as NH_3	0.3 lb NO_x/MMBtu

quench/cool down section, (2) a sulfur removal system, (3) a sulfur recovery section, and (4) an ammonia removal/sour water stripper. The thermodynamic losses associated with each of these sections was identified, either as a loss in sensible heat which otherwise could be used to generate power or as a direct requirement for fuel for combustion. The costs of sulfur removal were also calculated on the basis of $/lb S removed. The comparison of these cleanup system characteristics are given in Table 67.

Low-Temperature Systems

The coupling of the Selexol process with the BuMines-Type and BCR-type gasifier presents an opportunity to define the effect of low-temperature cleanup on performance since there is nearly twice the sensible heat in the BCR gas stream to be removed before scrubbing. Yet the net effect as shown in Table 67 is nearly the same, i.e., a 2.51 percent heat loss for the BuMines versus a 2.[illegible] percent loss for the BCR. The mechanisms of loss, however, are different enough to require discussion.

In the BuMines system, nearly 10.5 percent of the total fuel gas heating value is sensible heat; for the BCR, slightly over 19 percent. Because of the tar scrub required by the BuMines gasifier, regeneration of the clean fuel gas is possible to only [illegible]5 F, a process which recovers about 21 percent of the heat (2.19 percent of the total gas stream energy). The BCR gas stream allows regeneration to 1000 F, thereby recovering 51.2 percent (9.78 percent of total). Slightly over [illegible] percent of the BuMines gas stream sensible heat is used in the Selexol system versus 14.5 percent in the BCR stream (the COS in the BCR stream requires a higher solvent flow rate). Process heat furnished by the fuel gas sensible heat for ammonia removal/sour water stripping is 49 percent for the BuMines, [illegible] percent for the BCR.

In the sulfur recovery step, additional clean fuel is consumed in the Claus process. For the BCR-based process, 1.[illegible]4 percent of the total fuel energy is consumed; for the BuMines-based process, 0.71 percent. The higher requirement of the BCR process is due mainly to the more dilute gases resulting from the higher solvent recirculation rates needed to remove the COS.

The auxiliaries associated with these systems use power which can be expressed as equivalent fuel energy. Once again, due to the higher COS content, power for solvent recirculation results in a use of nearly 3.8 percent total energy for the BCR versus the approximately 1.7 percent for the BuMines.

Overall, the energy consumption of sulfur removal/recovery associated with the BCR-type gasifier is 7.7[illegible] percent of the total fuel gas energy while that for the BuMines-type is 4.0[illegible] percent. By allowing a higher SO_2 emission

TABLE 67

COMPARISON OF CLEANUP SYSTEM CHARACTERISTICS

	Low-Temperature Systems		High Temperature Systems	
	BuMines Selexol	BCR Selexol	BuMines Iron Oxide	BCR Consol
Gasifier Outlet - Temperature - F	1000	1800	1000	1800
Cleanup Inlet Temperature - F	100	100	1000	1750
Fuel Temperature to Turbine - F	265	1000	1070	1700
Fuel Gas Energy Utilization - %				
Sensible Heat Available[1]	10.47	19.10	11.83	19.10
Used to Regenerate	2.19	9.78	-----	-----
Used in Sulfur Removal System	0.64	2.76	$(4.3)^2$	$(0.7)^2$
Used in Ammonia Removal System	5.13	4.20	-----	-----
Net Heat Loss	2.51	2.36	$(4.3)^2$	$(0.7)^2$
Chemical Fuel Value for Sulfur Recovery	0.71	1.64	7.5	-----
Cleanup System Auxiliary Fuel Equivalent	1.74	3.78	1.9	0.4
Total Cleanup System Losses - %	4.96	7.78	13.7	1.1
Cost of Sulfur Removed - $/lb	0.059	0.076	0.052	0.050

(1) All sensible heat referenced to 59 F

(2) Chemical HHV required to supply process heat

for the BCR-based system, the losses associated with the removal of COS could be essentially eliminated. This could result in an over all penalty of less than 4 percent with a subsequent overall system efficiency increase of around two points.

The cost of sulfur removal for the BCR-based system is also higher, $0.076/lb S versus $0.059 for the BuMines-based systems. This higher cost is attributed to: (a) higher Selexol system costs due to the COS, and (b) the more costly regenerative heat exchanger (see gas cooling costs in Table 54).

High-Temperature Systems

The BuMines-based high-temperature product gas contains 11.83 percent of the total fuel gas energy as sensible heat while the BCR-based gas stream is again 19.1 percent. There is no sensible heat loss due to cleanup and essentially all the sensible heat is used at the efficiency of the combined cycle. The sulfur removal system requires extraction of heat from the power cycle (or from an auxiliary fired-boiler) to provide some process heat. For the BuMines/iron oxide system this amounts to an equivalent of 4.3 percent of the total fuel gas stream energy; the BCR/Consol requires an equivalent of 0.7 percent. (Since the heat required is relatively low grade, it is more efficient to extract it from the power system after some work has been performed by the combustion/expansion of the fuel gas than by removing sensible heat from the gas prior to combustion. This process energy required has been prorated back to equivalent fuel gas energy as explained in Section 4.)

The transformation of the H_2S removed to elemental sulfur requires a very significant amount of energy in the BuMines/Iron Oxide Process; approximately 7.5 percent of the total fuel gas energy. By converting the fuel gas H_2S content to some form other than elemental sulfur, a very appreciable savings could be realized. For the purposes of this comparison study, however, only elemental sulfur has been considered.

The BCR/Consol system has no net requirement for process heat for sulfur recovery.

The total energy consumption penalty, including auxiliaries, associated with the two high-temperature cleanup systems are 13.7 percent for the BuMines/Iron Oxide and 1.1 percent for the BCR/Consol. The cost of sulfur removal for the two systems are nearly equal at $0.052/lb S for the iron oxide and $0.05/lb S for the Consol system.

Comments on Cleanup System Selection

When comparing the various low- and high-temperature cleanup systems it is necessary not only to look at the energy penalties associated with each, but also to consider the net effect on overall cycle performance. For example, the two first-generation systems based upon the BuMines gasifier and the 2200 F COGAS system have essentially equal overall efficiencies (see Table 45) yet the penalties associated with sulfur removal indicate that those for high-temperature systems are roughly three times as great as for the low-temperature system. When the systems are integrated, however, the net result is that the sensible heat in the high-temperature systems offset nearly all of these losses so that the overall effect is that the integrated system with high-temperature cleanup is approximately 2 percent more efficient (0.6 efficiency points) than with the low-temperature cleanup.

The second-generation integrated systems show more dramatically the performance advantages of the high-temperature cleanup system, due mostly to more efficient utilization of the very sizable sensible heat content of the BCR product gas; i.e., approximately 9.5 percent more of the total gas energy is used at combined-cycle efficiency rather than steam cycle or process efficiency. This, coupled with the much lower process heat/auxiliary requirement, results in a 17 percent (6 points) performance advantage. It should be noted that a similar advantage (approximately [illegible] points) is also available if the advanced gasifier is coupled with a first-generation COGAS system (see Table 45).

As would be expected, the economics are more favorable for the generally less complex high-temperature systems. The absence of heat-exchangers and large-scale pumps and compressors for process work more than offsets the additional cost for high-temperature materials.

The foregoing must be tempered, however, by the realization that both the performance and cost estimates for the low-temperature system are based upon extrapolations of data for actual sour gas treatment while those for the high-temperature systems are based upon extrapolations from bench-scale or small pilot facilities. In addition, the emissions of NO_x and particulates from systems using the high-temperature processes are potentially unacceptable.

Nonetheless, the results of the analyses performed during this study indicate enough potential advantages for the high-temperature cleanup systems that further development in this area should receive every consideration.

TECHNOLOGY REQUIRED TO OBTAIN FUEL SPECIFICATION GOALS

The suggested fuel specifications given in Table 66 were arrived at through consideration of environmental aspects (sulfur and nitrogen components) and turbine requirements (alkali metals and particulates). The following paragraphs discuss these suggested fuel specification goals and their implication on the technology required to achieve them.

Technology for Sulfur Removal

The suggested fuel sulfur specification (0.05 Mol percent or 500 ppm) can be achieved in nearly all of the currently available low-temperature cleanup systems. These systems are capable of cleaning to 100 ppm or lower and, in fact, the technical and economic differences between a 500 ppm and 100 ppm limit are small.

Several of the high-temperature cleanup systems, such as those based upon iron oxide can clean to the 100 ppm level. However, the half-calcined dolomite system must have its operating conditions altered to achieve the 500 ppm goal. Since the operating temperature is fixed by the heat balance around the desulfurizer, the temperature differential of about 50 F needed to reduce the concentration from [illegible] to 500 ppm residual sulfur can only be obtained by raising the gasifier operating temperature. Alternatively, since the residual sulfur content is inversely proportional to total pressure, (see Figure 47 shown previously) a 10 percent reduction in operating pressure would result in achieving the goal of 500 ppm sulfur in the treated gas.

The foregoing does not require technology changes, rather it requires consideration of altered operating conditions. However, the use of high-temperature fuel gas would require changes in the technology of the gas turbine. Currently, the control system of the gas turbine is based upon the metering of fuel into the burner. Machines in operation today are capable of handling fuel to 400 F in these controls. Research is being carried on to develop a fuel control system which will work to over [illegible] F. It appears that considerable attention must be paid to developing either a fuel control which operates to 1750 F or above, or to developing new integrated system controls which would use the gasifier input as the control element. Because of the thermal inertia represented by the gasifier, it is not apparent that control from in front of the gasifier will give satisfactory performance under utility operating conditions.

Technology For Fuel Nitrogen Removal

As with the sulfur removal, it is not necessary to develop new technology to remove nitrogen compounds, particularly ammonia, from the effluents coming from the gasifier and low-temperature cleanup process. The low-temperature scrubbing systems utilize water wash which will remove most of the ammonia. In addition, the sulfur removal system will absorb ammonia, Selexol, for example is more selective to ammonia removal than to carbon dioxide. The process is most selective to HCN, another potential fuel nitrogen compound. Thus, fuel nitrogen content of low-Btu fuel gas from these scrubbing systems is quite low and does not appear to contribute significantly to NO_x formations.

Unfortunately, the high-temperature processes including the iron oxide or dolomite-based sulfur removal systems do not remove any of the ammonia or other nitrogen-based compounds from the fuel gas. As can be seen in Table 57, the fuel nitrogen does contribute to overall NO_x production. Catalytic decomposition of ammonia to N_2 and H_2 has the potential to reduce ammonia content to perhaps 200 ppm for the BCR/CONSOL. However, a catalysis system capable of working under the conditions assumed present in the high-temperature stream, e.g. 1600 F to 1700 F at approximately 500 psi has yet to be identified. This method does show the only promise and should be further investigated.

Technology for Particulate Removal

The low-temperature scrubbing systems are capable of removing particles to the level of cleanliness beyond those specified in Table 66. Thus, little problem is expected to arise from fuel introduced particles with this process. Once again, the high-temperature system would be subject to particulate based problems.

A review of the literature indicated very little work had been done at the temperatures of interest. Some work has been done on developing electrostatic precipitators for operation up to 1500 F although these systems have not had significant commerical application. (69) Of greater promise, is the development of high-temperature filtration systems through beds of granular solids such as the panel bed sand filters which claim 99 percent fly ash removal efficiencies. (70, 71) This concept has not yet been demonstrated on a commercial scale.

A previous study (65) had investigated various devices for removing particles in jet engine exhausts at temperature to 1100 F. Dry cyclones, fabric and ceramic filters and electro-static precipitators were investigated as were removal methods which were still in an early stage of development such as acoustical and magnetic separators. None of these were attractive for the engine exhaust application, either because of the very high flow rates involved or of the high-temperature. These methods were reviewed and it was determined that cyclones and ceramic filters could handle both the flow and the temperatures associated with the high-temperature fuel gas stream although their removal effectiveness was not necessarily adequate.

It was pointed out in a previous study (3) that the rotary stream dust collector is a form of cyclone produced by the Aerodyne Corporation and known as a tornado. This technology is well known in Europe. Aerodyne claims essentially 99 percent collection for particles of 5μ and about 70 percent for 1μ.

Unfortunately, the size distribution of the effluent particles from a gasifier are not known. For purposes of this study, it was assumed that the dust carry over for the BOM gasifier, roughly 10 percent of the coal ash, would be present in the off gas from both the BOM and BCR gasifiers. It was further assumed that the dust carryover resembled fly ash and demonstrated weight fractions shown in Table 68.

The ability of the turbine to withstand particulates is a function of particle size and composition. Unfortunately, there is little documentation other than that already cited.(67, 68) Measurements of particle sizes in gas turbine exhausts indicate that most (95 percent-plus) are less than 1μ with a median size of 0.1 (65). This would indicate that particles above 1μ are fractured either by the high-turbulence within the turbine, or by collsions with other particles or turbine parts. Since fly ash is a fluffy friable material, easily broken, a 2μ upper limit was set (this is less than the 6μ limit suggested in the Australian work (67), but would result in a conservative estimate of capture effectiveness).

Based upon the foregoing, it appears that a cyclone-based system using an Aerodyne-type final stage capable of removing 99 percent above 2μ (equivalent to < 0.04 lb/MMBtu) would be suitable since it is well below the EPA turbine limit of Table 64. However, this value (0.05 gr/ft³) is about a factor of five higher than the gas turbine fuel specifications for methane (Table 65) and a factor of 40 higher than the suggested specification for low-Btu fuel (Table 66). Thus, the ability to meet and actually surpass the EPA regulation does not, in this case, satisfy the governing requirement, i.e., the capability of the turbine to withstand particulate ingestion. A review of the very recent literature and discussions with Pratt & Whitney Aircraft revealed only one new bit of data on particulate loading, the estimate by Westinghouse (72) that, in the pressurized fluid-bed effluent, a loading of 0.15 gr/ft³ would be suitable for the turbine. This is roughly 70 percent higher than the EPA limit. The turbine for the fluid-bed application operates at relatively low temperature and these values may not be applicable in the higher temperature cases. Thus, it is probable that more sophisticated particulate removal devices would be required.

Table 68

SIZE DISTRIBUTION OF FLYASH FROM BOILERS

(cumulative weight % above a given size)

Size, μ	Illinois No. 6	Wyoming
595 (30 mesh)	0.0	0.1
297 (50 mesh)	0.2	0.3
149 (100 mesh)	1.6	1.8
74 (200 mesh)	10.9	11.2
40	39.0	30.[illegible]
[illegible]	54.0	49.0
10	82.0	76.0
[illegible]	94.0	89.0
[illegible]	98.0	95.0
0	100	100

RECOMMENDATIONS

The foregoing study has shown that the potential thermodynamic advantages of high-temperature fuel gas cleanup are reduced by the operating characteristics and requirements of the cleanup systems and, further, that the potential emission of NO_x due to fuel nitrogen is unacceptably high compared to the EPA standards. Nonetheless, the overall efficiencies of those integrated power stations using high-temperature cleanup are sufficiently higher (20 percent) that efforts aimed towards early commercialization are well warranted. Specific recommendations for further effort are given in the following paragraphs.

Gasification

The BOM gasifier and the BCR gasifier were selected as representative of two generic types of gasifiers, a low-temperature type having tars in the effluent, and a high-temperature type having tar-free off gas. Actually, the tar content has little effect on the cleanup system as it can be rather easily removed from the off gas prior to the low-temperature sulfur cleanup step and simply passes through the high-temperature cleanup in the vapor phase. As it turns out, the ammonia (fuel nitrogen) content appears to be the major problem area. While easily removed in the low-temperature system, the subsequent ammonia stripping and recovery are costly both in money and energy. High-temperature removal of ammonia, while potentially possible, has not yet been demonstrated in a form similar to that which would be required for use in conjunction with pressurized gasification.

It would be warranted to examine a gasifier which does not produce ammonia, i.e., a very high-temperature system which would essentially crack fuel nitrogen compounds to nitrogen, hydrogen and carbon. One such gasifier presently operating on coal is the Koppers-Totzek entrained-flow gasifier.

The gasification of high-sulfur residual oil or of other high-sulfur petroleum based products such as petroleum coke has been considered as an alternative fuel supply system to power stations currently burning oil. While still of interest in other areas of the world, particularly the USSR and Japan, this process is currently in limbo in this country because of the high costs of oil. Yet, future energy scenarios such as those considering the impact of Alaskan oil on the mainland West Coast could provide significant supplies of residual for potential use in utility powerplants.

The present study has been limited to gasifier operation at a single point because of the lack of hard, reliable data on gasifier operation. Since it will be several years before any large-scale gasifier operation will occur, other than the fixed-bed type, it would be advantageous to develop the capability to estimate the relationships of the operating conditions such as pressure, temperature, air and steam flow rates to the efficiency of the gasifier and to the makeup of the fuel gas. This could be accomplished for entrained-flow type gasifiers through the use of a computer model based upon quasi-equilibrium chemical relationships.

It is recommended that further study be made of an integrated power system based upon the Koppers-Totzek gasifier.

Based upon the foregoing comments, the following recommendations are set forth:

It is recommended that a study be performed to determine the technical, economic feasibility and environmental impact of an integrated residual oil, low- and high-temperature cleanup and combined-cycle power system.

It is recommended that a computer model of the coal gasification process be developed.

Low-Temperature Cleanup Systems

The low-temperature cleanup systems displayed the ability to remove H_2S from the fuel gas to levels of 100 ppm or below. Fuel nitrogen compounds were also removed as were particulates. However, the utility loads (process steam, electricity) were such that overall system efficiency was compromised more than had been suggested by previous, but perhaps, superficial studies. These inefficiencies are due in part to the process requirements for the aforementioned ammonia stripping as well as for sulfur absorber solvent stripping. Several other additional process heat requirements for sulfur reclamation are present.

It is recommended that the low-temperature scrubbing processes be reviewed to determine the effect on overall system performance of modifications to the process.

High-Temperature Cleanup Systems

The type of inefficiencies associated with the low-temperature process are not present with the high-temperature processes. This may be in part due to the early state of development of these processes with its attendant lack of operational data. The major challenges associated with the high-temperature cleanup are the removal of fuel nitrogen compounds and particulates. The latter has a broader technological base and has several potential solutions undergoing early phases of testing. Ammonia removal does not have an equivalent base on which to build.

It is recommended that programs aimed at removing particles of 2μ to 5μ from high-temperature gas streams be initiated or strengthened.

It is recommended that methods of removing ammonia from high-temperature gas streams be identified and their technical and economic feasibility be established.

Advanced Power Systems

The advanced power systems used in the present study are based upon two levels of technology. The first level is that which is being applied to turbomachinery currently being developed for introduction in commercial service in the 1[illegible] - [illegible] time frame. The basis for the performance of the first-generation turbomachinery is the use of air cooling in the hot section of the turbine, a concept well proven in current industrial and aircraft applications.

The performance for the second generation turbomachinery is based upon the contractor's assessment of the advancements that will be made in advanced cooling technology and materials.

These advances will occur at rates directly proportional to the R & D effort expended, i.e., the funding made available to bring the concepts to the commercialization phase.

It is recommended that a program be initiated to demonstrate high-temperature gas turbines which would be capable of attaining 50 percent efficiency in the combined-cycle mode using clean fuel.

Environmental Impact

The present study gives quantitative estimates for the emissions of SO_2, NO_x, and particulates. Of these, only the SO_2 emissions are based upon actual data, i.e., measurements of H_2S and COS in fuel gas streams. The estimates of NO_x emissions are based upon models using a mixture of theoretical and empirical data. The latter are extrapolations of a small amount of low-pressure combustion data on low-Btu gas from an entrained-flow gasifier. The estimates of particulates are based on the assumption of similarity of fly ash emissions from boilers and gasifiers.

There are potentially several methods other than cyclones to remove particles from high-temperature streams. The use of panel-bed filters has been mentioned previously. While none of these devices has been operated at the conditions considered typical for the gasifier off gas, some small scale tests have been conducted. In particular, a sand filter is being developed by the Combustion Power Corporation under partial EPA sponsorship.(73) The test data available indicates a potential for removal of < 2μ particles, but several problems remain to be solved before a successful pilot plant could be operated, much less a commerical-sized installation.

During the course of investigating the high-temperature particulate removal devices, it was discovered that the Stone & Webster Co. was doing essentially the same type of investigation under the auspices of the Electric Power Research Institute (EPRI). A discussion was held with Stone & Webster in order to establish the status of their work and to determine if any potentially attractive systems had been overlooked. It was ascertained that the two studies had reached roughly the same conclusions, i.e., a complete lack of actual, reliable data on the particulate carryover from gasifiers and also on the ability of the turbine to withstand particulate ingestion. In addition, it was learned that metallic filters capable of 99.5 percent-plus removal of particles down to 0.5μ had been developed by the Brunswick Corporation. These materials are used as seals in high-temperature (2200 F) gas turbine engines. This material, called Brunsmet, has been tested in an 8000 cfm capacity pilot facility and has operated at conditions indicating an advantage over normal bag house materials by a factor of 25 (25 times more air can be handled per unit area). Costs for this system are not available but Brunswick Corporation claims an economic advantage over conventional materials.

Stone and Webster indicated that small-scale tests of a panel bed filter at the BOM in Morgantown were favorable, although efficiencies were of the order of 99 percent rather than 99.9 percent. Problems with the test apparatus other than the actual filter could account for the discrepancy according to Stone and Webster.

From the above, it can be concluded that the technology for particulate removal down to 2μ does exist although it has not yet been demonstrated on a scale suitable for subsequent commercialization. It is recommended that target-scale testing be carried out at conditions typical of gasification systems, e.g., 300-500 psig, 1400-1700 F and that design studies for mounting and support systems for holding panel beds or metal mesh systems be initiated.

It is recommended that further development of the computer model of NO_x formation arising from the use of low-Btu fuel gas in compact combustors be carried out using a broader base of empirical data from various combustor programs currently being performed on low-Btu gas.

It is recommended that measurements be made of particulate carryover from existing gasifiers in order that more realistic estimates of particulate cleanup can be made, and that more complete information on the total environmental impact of the total systems be obtained.

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APPENDIX A

PERFORMANCE EVALUATION PROGRAM

System performance calculations and gas turbine design and sizing estimates were made using programs developed by the contractor for use in corporate sponsored programs. These are described in this appendix.

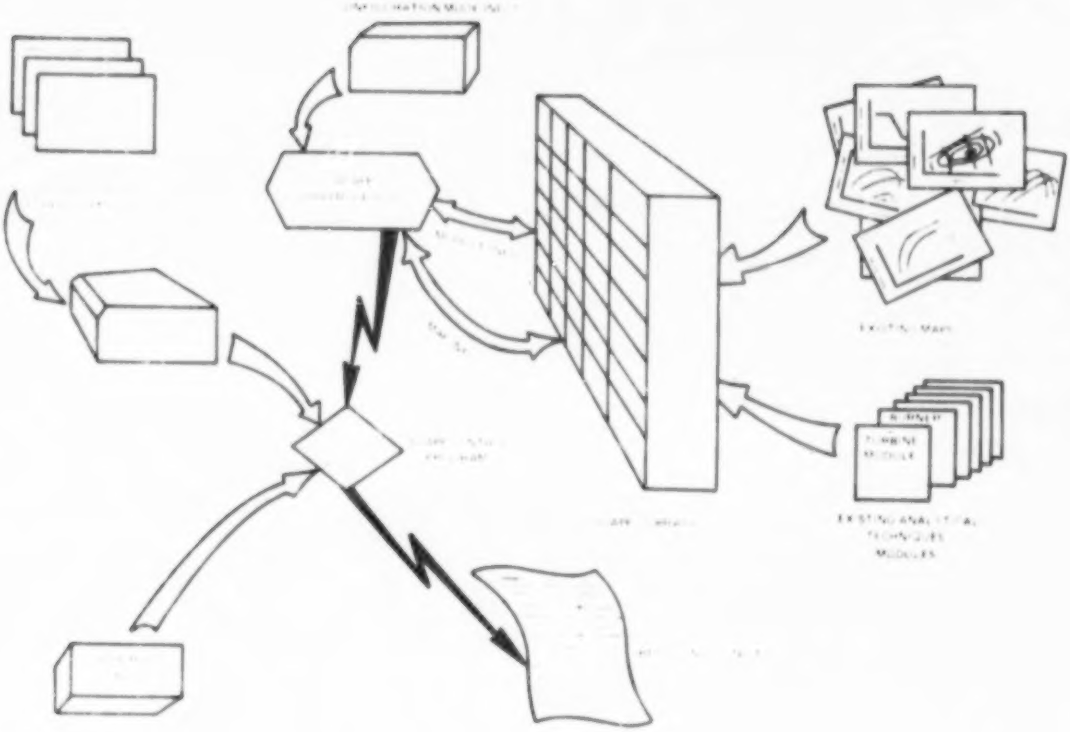
State-of-the-Art Performance Program (SOAPP)

The SOAPP Program is a revolutionary and sophisticated computational system which can be used to analyze almost any complex power system configuration consisting of a large number of components. SOAPP could also be used to estimate the design, economic, and environmental characteristics of advanced-cycle power systems if desired. The SOAPP system is based on a completely modularized representation of system components. Modularization permits considerable versatility in selecting power system configurations to be analyzed and allows continual update of the system as improved revisions of each module became available. Many revisions of these preprogrammed modules with differing degrees of complexity are stored in an extensive SOAPP library.

The central and unique feature of SOAPP is a preprocessor or precompiler which establishes the sequential logic required for system calculations and performs all the detailed programming work necessary for any configuration. Figure 68 presents a pictoral representation of how SOAPP operates. The desired configuration is specified using a simple alphanumeric code, along with specific module and performance map identification codes. The preprocessor then selects the corresponding module and map routines from the library and writes the main control program in FORTRAN which will include all the mathematical logic necessary to provide complete mass and energy balance for the desired configuration. The control program performs all calculations in proper sequence and accounts for iterative balances of design requirements, in and out bleed or extraction streams, external schedules and controls, and transfer of data between modules.

FIG. 68

OPERATION OF SOAPP SYSTEM


BURNER
TURBINE MODULE
EXISTING MAPS
EXISTING ANALYTICAL TECHNIQUES MODULES

The configuration flexibility described above was the primary goal of SOAPP. Other advantages of SOAPP involve user convenience features such as automatic cycling of run data, data input/output flexibility, and automatic balancing of constraints imposed on parameters at intermediate locations in the system.

Gas Turbine Flowpath Design

A high-speed digital computer program was developed under Corporate sponsorship to facilitate parametric performance and sizing studies of gas turbine designs incorporating simple cycles, intercooling, regeneration, and multiple shafts. The program also provides a realistic assessment of turbine cooling flow penalty effects on gas turbine thermal efficiency and specific output. Once the primary independent variables (e.g., material technology, turbine inlet gas temperature, compressor pressure ratio, and airflow rate) have been specified, the program computes the gas flowpath, pertinent dimensions, number of compressor and turbine stages, and turbine cooling flow requirements.

Gas Turbine Cooling Flow Calculations

One of the major problems facing gas turbine manufacturers today is proper design of cooling systems to restrict metal temperatures to levels low enough to ensure long life and minimize maintenance problems. Several computer programs have been developed which can be used to estimate the amount of cooling airflow required in terms of local hot gas temperature, maximum allowable metal temperature, and the geometry of the cooling flow passages. One of these programs was modified by the contractor to provide output which is compatible with SOAPP, especially when low-Btu gasified fuels are used in the engine.

APPENDIX B

GAS TURBINE COMBUSTOR POLLUTION EMISSION MODEL

This Appendix contains a description of a gas turbine combustor model developed by United Technologies Corporation for use in predicting exhaust pollutant emissions. The original version of this model(74) divided the combustor into three zones in which liquid droplet burning was modeled. This original model was subsequently revised(75-80) to incorporate improved simulation techniques for the principal aerodynamic and combustion processes in the complex combustor flowfield. In this improved version, aerodynamic properties of the flow field and concentrations of carbon monoxide, total unburned hydrocarbons, and nitric oxide were predicted as a function of radial and axial position within the combustor for a variety of current combustor configurations. The analysis directly considers details of the combustor geometry, the fuel injection system, and engine operating conditions.

The approach taken in the development of the original combustor model was (1) to formulate mathematical treatments for the principal physical and chemical mechanisms that influence the combustion process, and (2) to integrate these mechanisms through a sequence of thermodynamic states obtained from the coupling of these mechanisms with the physical combustor flowfield equations to provide the gas temperature, flow velocity, and chemical species concentrations as a function of position within the combustor which, in turn, influence subsequent combustion. The principal elements of the analysis are a combustor internal flowfield model, a physical combustion model, a treatment of hydrocarbon-air chemical kinetics, and a NO_x kinetics model.

The original P&WA liquid fuel combustor model was subsequently modified by the United Technologies Research Center, at Corporate expense, to permit the model to be used for low-Btu gasified fuel. The combustor internal flowfield model and the NO_x kinetics model were basically unchanged. The physical combustion and hydrocarbon-air chemical kinetics models were revised to simulate the combustion and mixing characteristics of low-Btu fuel gas. The remainder of this appendix deals with the highlights of the low-Btu combustor model with emphasis on revisions made to the original P&WA combustor model.

Internal Flowfield Model

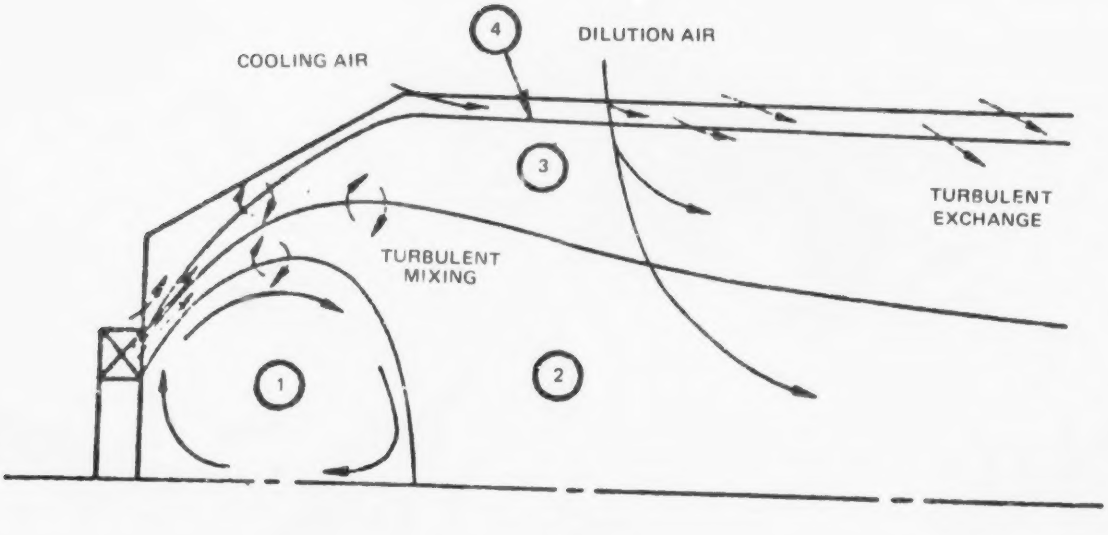
The combustor flowfield model defines the physical system on which the gas dynamic and combustion rate calculations are based. The experimentally determined internal flowfield for a conventional swirl-stabilized, can-type combustor is shown in Figure 69. The flowfield is seen to include a region of highly turbulent, reversed flow in the front of the combustor, surrounded by a region of relatively uniform downstream flow. The forward region, including the recirculating zone, is designated the primary zone and the downstream region, the secondary or dilution zone. The primary zone serves the purpose of stabilizing the combustion process; combustion is largely completed within this zone. The mixture of high-temperature combustion products and reactants leaving the primary zone continues to burn and subsequently is mixed with dilution air in the secondary zone to provide a suitable temperature profile for entrance to the turbine.

The combustor flowfield model employed in the present analysis is shown schematically in Figure 70 for the case of a can-type combustor. The two-dimensional internal flowfield has been approximated by a set of coannular, one-dimensional reacting stream tubes. The recirculating zone boundary, enclosing region 1 of this figure, defines the location and size of a zero net flow, one-dimensional stream tube representing the recirculating flow. Air entering the front of the combustor is assigned to the main flow stream tubes on an equal basis. Downstream combustion and dilution jet air is apportioned to the stream tubes by means of a jet penetration and mixing model. All wall cooling air is assigned to the outer stream tube that begins at the first cooling air addition site. The airflow distribution to combustion and cooling holes is specified as model input. The stream tube boundaries are defined by inner and outer radii and are computed as dependent variables. The outermost stream tube is bounded by the location of the chamber wall, which is also provided as input.

Physical Combustion Model

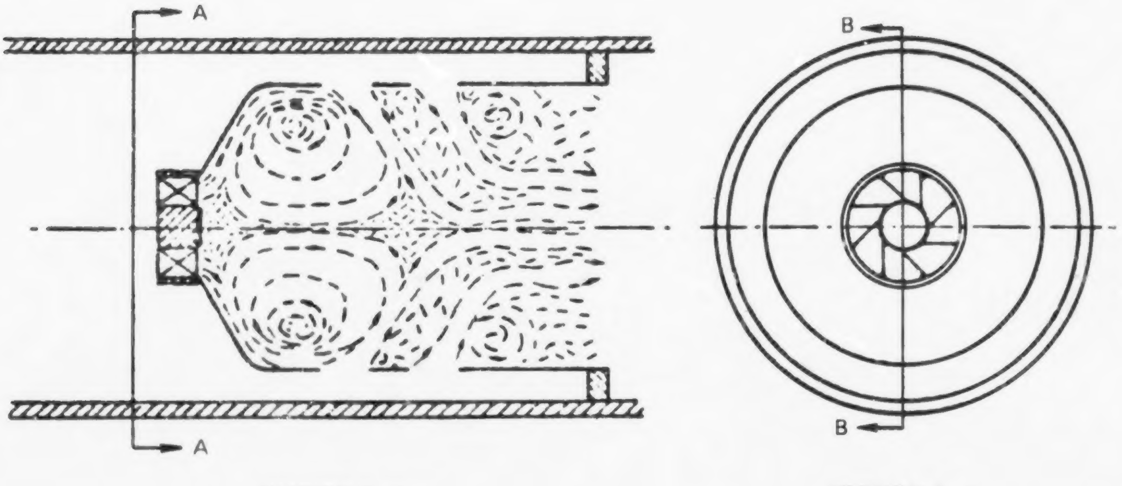
The combustion simulation used in the original P&WA liquid combustor model considered the complex oxidation of hydrocarbon fuel as occurring in three broad stages. The first stage provides for vaporization of the liquid fuel and breaking down the complex mixture of long chain hydrocarbons into light, unburned, partially oxidized hydrocarbons. The subsequent sequence of reactions, comprising the second stage of combustion, includes the principal exothermic reactions and produces large amounts of H_2O and CO. The final stage of combustion is characterized by the conversion of CO to CO_2.

SWIRL–STABILIZED CAN FLOW FIELD MODEL


4
COOLING AIR
DILUTION AIR
3
TURBULENT EXCHANGE
TURBULENT MIXING
1
2

N03–80–3

PRIMARY–ZONE FLOW PATTERN OBSERVED IN A CAN-TYPE COMBUSTOR



SECTION B–B

SECTION A–A

An essential feature of the liquid combustion model described above is the rate equations which govern breakdown of the long chain hydrocarbons. Because of this, the liquid combustion mechanism is not appropriate with fuel gas consisting of carbon monoxide, hydrogen, and methane. Hydrogen in the fuel gas will ignite and burn over a wide range of conditions as soon as it comes in contact with oxygen in heated air. The temperature rise from the hydrogen combustion would then cause the methane and carbon monoxide to start burning. The controlling parameter in fuel gas combustion, therefore, appears to be the time associated with mixing hydrogen in the fuel gas with oxygen in the air.

Turbulent mixing has been successfully simulated using eddy coefficient viscosity models for point-to-point diffusion in a large number of fluid dynamic models for which the difference equation grid was sufficiently small compared to a mean turbulent eddy size. These models have been successful both for energy and mass transfer. In the Pratt & Whitney combustor model the axial grid is very fine ($\Delta x = .05$ inch); however, the cross-axial grid is coarse. (The stream tube widths (see Figure 67) are often over an inch.) Therefore, the cross-axial mixing of hydrogen and air cannot be simulated by an eddy viscosity diffusion model. Also, an ignition delay mechanism, keyed to a specified temperature, would not work because temperature is averaged over a stream tube. Thus, a local hot region, e.g., bordering the recirculation zone, cannot be distinguished from the colder flow in a stream tube. The average temperature across the stream tube would be well below the ignition temperature even though ignition had occurred in part of the stream tube. An alternative model to eddy viscosity and ignition delay is needed to simulate the hydrogen-oxygen turbulent mixing process in each stream tube.

Equation 46 offers a simple model of the axial hydrogen-oxygen mixing in a stream tube which can be adapted to the Pratt & Whitney combustor model. The time rate of increase of burned vapor (fuel gas, including hydrogen, plus air) is assumed proportional to the product of the molar concentrations of unburned fuel vapor and oxygen. This assumption is analogous to the probability of collision theory customarily used in statistical thermodynamics(81).

The difference in the argument is that the time scale chosen corresponds to a typical turbulent mixing time rather than a thermodynamic time scale related to average particle velocities. Thus:

$$\frac{d}{dt^*}[WBV]^* = [WUV]^*[O_2]^* \qquad (45)$$

---

* denotes nondimensional quantities - that is, quantities scaled so as to be approximately unity.

where:

$$[WBV]^* = [WBV]/[WBV]\ \text{average}$$

$$[WUV]^* = [WUV]/[WUV]\ \text{average}$$

$$[O_2] = [O_2]/[O_2]\ \text{average}$$

$$t^* = t/T$$

and:

$[WBV]$ = moles of (burned fuel gas + air)/unit volume

$[WUV]$ = moles of (unburned fuel gas + air)/unit volume

$[O_2]$ = moles of oxygen/unit volume

T = mixing time = $L/\sqrt{u'^2}$

L = mixing length

$\sqrt{u'^2}$ = r.m.s. turbulent velocity fluctuation

The mixing time formula given above is commonly used in fluid dynamic analyses(82-84). Choosing representative values for the dimensional parameters identified above leads to a more useful form of Equation 45 given below:

$$\frac{d[WBV]}{dt} = R\ [WUV]\ [O_2] \qquad (46)$$

where:

$$R = 10^9\ \text{cm}^3\text{-mole}^{-1}\text{-sec}^{-1}$$

It was found that the simple mixing model embodied in Equation 46 yielded primary zone lengths which were always in the expected 2 to 4 inch range for the FT4 combustor over a broad range of fuels and operating conditions. This gives confidence that Equation 46 does simulate hydrogen-oxygen turbulent mixing adequately for this combustor model.

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* denotes nondimensional quantities - that is, quantities scaled so as to be approximately unity.

NO_x Kinetics Model

Nitric oxide (NO) is generally the most prevalent oxide of nitrogen observed in the exhaust gas from gas turbine engine combustors. Small amounts of nitrogen dioxide (NO_2), however, are also observed and are thought to be formed by the oxidation of the previously formed nitric oxide. Consequently, an evaluation of nitric oxide production within the combustor flowfield essentially constitutes a complete assessment of gas turbine engine combustor-generated oxides of nitrogen.

Several researchers have been successful in describing the mechanisms involved in the formation and depletion of the oxides of nitrogen. Of these, however, Lavorie, et al(35) identified specific reaction mechanisms involving the principal reactions responsible for the formation and depletion of nitric oxide. Consequently, the mechanisms of Lavorie (see Table 69) were used in the oxides of nitrogen concentration prediction system.

Equilibrium concentrations of the species O, O_2, H, OH, and N_2, corrected to the local carbon monoxide level, if in excess of equilibrium, are assumed in applying the reactions shown in Table 69. The deviation of the local CO concentration from its corresponding equilibrium values was assumed to be a measure of the proximity to thermodynamic equilibrium of all other species involved. Since relaxation times of the species N and N_2O are typically several orders of magnitude less than that for NO, steady-state concentrations for N and N_2O may also be assumed. The expression for the rate of NO reaction thus reduces to (following Lavorie):

$$\frac{d[NO]}{dt} = 2(1 - \alpha^2)\left[\frac{R_1}{1 + \alpha K_1} + \frac{R_6}{1 + K_2}\right] \qquad (47)$$

where:

$$K_1 = \frac{R_1}{R_2 + R_3}$$

$$K_2 = \frac{R_6}{R_4 + R_5}$$

$$\alpha = [NO]/[NO]_e$$

R_i is the one-way rate of the i^{th} reaction; for instance:

$$R_1 = K_{11} [N]_e [NO]_e \tag{48}$$

and the subscript "e" denotes equilibrium concentration.

Reactions 1 through 3 in Table 69 correspond to the Zeldovich(36) mechanism as modified by addition of the reaction between the species N and OH. The reaction rates shown have been updated from the originally reported work using more recent data(37-39).

Table 69

NITRIC OXIDE FORMATION KINETICS RELATIONSHIPS

Reaction		Forward Rate Constant cm3/mole-sec
Number	Equation	
1	$N_2 + O = NO + N$	$R_1 = k_{11} = 1.35 \times 10^{14}e-37,500/T$
2	$N + O_2 = NO + O$	$R_2 = k_{12} = 6.4 \times 10^{9}Te^{-3},125/T$
3	$N + OH = NO + H$	$R_3 = k_{13} = 7 \times 10^{11}$
4	$H + N_2O = N_2 + OH$	$R_4 = k_{14} = 4 \times 10^{13}e-6,000/T$
5	$O + N_2O = O_2 + N_2$	$R_5 = k_{15} = 5 \times 10^{13}e-14,000/T$
6	$O + N_2O = NO + NO$	$R_6 = k_{16} = 2.5 \times 10^{13}e-13,450/T$

APPENDIX C

POWERPLANT COST ANALYSIS

Previous studys (3, 4) demonstrated the need for a capability to provide cost estimates for future power systems that not only would enable power stations with different operating characteristics be considered, but also would allow sensitivity studies to be undertaken for differing technical characteristics of the selected systems. As a result, members of the technical staff at UTRC, in cooperation with cost estimators at Burns & Roe, Inc., developed a set of correlations based on histrocial data which enable the total installed cost of COGAS power stations to be estimated in-house with a high degree of confidence. The procedure consists of a set of equations, grouped by FPC account numbers, which can be used to calculate the installed costs of major power station equipment.

Minor component equipment items are combined in logical groups, the costs of which are also estimated from correlation equations. The calculation procedure requires as input data which are normally calculated as part of a routine technical analysis for electric power stations. Examples of such data include: gross and net station output power; steam turbine output power; gas turbine output power; cooling water flow rates; subcomponent input power requirements; and station efficiency. Because the cost correlations have been developed in sufficient detail, the need to resort to expensive preliminary system layout drawings is eliminated.

As part of the overall analysis, the costs of a few major components are estimated by methods other than the correlations described since these alternative methods make it possible to determine these specific equipment costs in even greater detail. Two such components are the large, industrial gas turbines and waste heat boilers, the costs of which are calculated using sophisticated computer program analyses proprietary to UTRC.

Furthermore, the costs of large, "standard" items such as steam turbo-generators and electric generators are obtained directly from published catalog data, which are corrected by appropriate price multiplier factors to maintain consistency with the latest industry experience.

A detailed list of all system components whose prices are capable of being estimated directly is presented in Table 70 for all applicable FPC power station account categories. Not all items are used for each COGAS system considered since often there are specific characteristics of each station design which distinguish it from other stations. In the UTRC procedure, allowances are also made for such items as station start-up, temporary buildings, transportation during construction, special tools required during a particular power station contract, engineering, contingency, escalation (except where included in the cost of selected major system components), and interest during construction. Examples of these are shown to the right of each category for a 1000 Mw, nominally-rated COGAS power station design.

The calculation includes allowances for profit, insurance, and fringe benefits directly in the cost of each component or group of components. This approach was followed: (1) because it protects the confidentiality of program profitability at the A&E firm; and (2) because changes in specific station details, once a set of estimates are completed, could be made easily without the necessity of changing an entire analysis set. This approach has been successfully demonstrated (4) at UTRC. Its only restriction is that it cannot estimate directly the effect geographic location has on overall power station costs. However, this problem can be rectified easily by using other cost analysis techniques(50). This is done by equating the sum of the costs of groups of components identified in the Ref. 50 analysis with the sum of costs of the same group of components determined in the UTRC analysis. (UTRC estimates can be grouped into the general categories of materials and equipment, labor, fringe benefits, supervision,construction tooling, engineering, contingency, insurance, and profit.) Segregating power station costs in this manner then allows the geographic location problem to be addressed directly through use of the location (labor) factors for major U.S. cities.

The following example illustrates the relationship between the UTRC procedure and that of Ref. 50. On the right side of Table 70 are presented costs of the noted components for an illustrative COGAS power station. These costs, expressed in 1974 dollar values, have been estimated from the UTRC correlations described in this appendix. The costs are presented by account number in Table 70; to the total have been added allowances for other expenses during construction, contingency, engineering and supervision, escalation, and interest (at a rate of 10 percent per annum) during a four year construction period. For this example, total turn-key power station cost for an Eastern Seaboard (New York/New Jersey) location is estimated to be slightly more than

Table 70

SPECIFIC COMPONENTS WHOSE COSTS ARE CALCULATED IN POWER STATION ANALYSIS

FPC No. 312: Boiler Plant Equipment

Waste Heat Boiler	$27,591,900
Boiler Feed Pumps	339,390
Water Demineralizer	4,630,590
Condensate Storage Tank	29,940
Process Steam Heat Exchanger	In Acct 343
Miscellaneous Pumps	70,525
Piping and Pipe Insulation	3,344,660
Computer and Associated Boiler Plant Controls	308,025
Total, Account 312:	$36,448,100

FPC No. 314: Turbine-Generator

Steam Turbogenerator	$ 9,833,270
Pedistal	In Acct 341
Condenser	1,212,180
Condensate Vacuum Pumps and Motor	134.795
Condensate Pump and Motor	274,780
Cooling Tower	6,618,750
Water Circulating Pump, Valves, and Expansion Joints	1,146,280
Make-Up Structure: Screen and Pump	
Chlorination Equipment	
Lube Oil Purifier	In Acct 341
Total, Account 314:	$19,217,055

FPC No. 341: Site and Peripheral Structures

Site Preparation	$ 848,250
Administration Building	563,830
Turbogenerator Building	3,948,000
Condensate Polishing System	837,065
Tank Form	1,236,100
Stack	362,000
Total, Account 341:	$ 7,795,245

Table 70 (Cont'd)

FPC No. 343: Gas Turbine

Gas Turbine	$17,643,500
Starter Motor	85,600
Torque Convertor	804,000
Lube Oil Purifier and Storage	224,000
Lube Oil Fire Protection	160,000
Turbine Air Coolers	320,000
Air Compressor Service and Instrumentation	140,000
Breeching	2,140,600
Expansion Joints	N.A.
Inlet Air Filters	555,760
Turbine Air Cooler Enclosure	N.A.
Emergency Cooling Water Tank, Pump and Piping	11,200
Fuel Oil Heaters and Pumps	123,200
Miscellaneous Pumps and Tanks	56,000
Control Panels	560,000
Fuel Piping and Pipe Insulation	1,436,900
Cooling Air Compressor and Motor	22,782,485
Total, Account 343:	$47,603,245
FPC No. 344: Electric Generator for Gas Turbine	$10,537,630
FPC No. 345: Accessory Electric Equipment	$ 8,768,750
FPC No. 346: Miscellaneous Power Plant Equipment	$ 359,040
FPC No. 353: Station Equipment	In Acct. No. 345

$200 million. The alternative procedure(50) provides the capability of estimating costs through the UTRC estimates defined as DIRECT CONSTRUCTION COSTS in Table 71. At the bottom of Table 71 is presented a breakdown by general cost category according to the general cost model(50) for two geographic areas: the noted Eastern Seaboard New York/New Jersey location; and a North Central (Illinois, Ohio, Pennsylvania) location. In the North Central total, the primary change in overall cost is due to the lower, mid-west labor rate based on labor factors(50). It can be seen that for the eastern location, DIRECT CONSTRUCTION COSTS are identical to the same total detailed in the upper half of this table. Because of the different approaches represented by the two models, it is difficult to make direct cost comparisons. However, both models make reasonable estimates of the desired station costs, and therefore focus on the end results is recommended rather than on the detail differences which often are the results of judgmental estimates.

The UTRC calculation procedure has been written in such a manner that corrections by single or grouped component(s) can be made as costs vary over time and with inflation rate. The Handy-Whitman Index of Public Utility Construction Costs(83), a regularly updated set of construction cost indexes recognized throughout the utility industry, has been applied where necessary to all cost equations in the UTRC procedure. As noted, all cost values are presented in terms of constant 1974 dollars. However, a projected future escalation rate must be selected and applied. In such cases, it is advised that all efforts proceed with extreme caution because of the volatility associated with projections in the unsettled financial environment confronting utilities and manufacturing organizations in the U.S.

Table 71

ILLUSTRATIVE EXAMPLE OF ELECTRIC POWER STATION COST ESTIMATING

Station Location: New York/New Jersey Area
Detailed Background Data in Table 70

312	$36,448,100
314	19,217,055
341	7,795,245
343	47,603,245
344	10,537,630
345	8,768,750
346	359,040
	$130,729,065
Other Expenses	2,614,580
Contingency	10,667,490
Engineering and Supervision	20,001,550
DIRECT CONSTRUCTION COST	$164,012,685
Escalation	18,441,775
Interest During Construction	18,245,450
Total (Turn Key) Cost	$200,699,910

EPA General Cost Model Approach

	NY/NJ Labor Factor: 1.875) Cost, 1974 $	Ill., Ohio, Pa (Labor Factor: 1.633) Cost, 1974 $
Materials + Equipment	$81,806,950	$81,806,950
Labor, Fringe, Supervision, Admin.	35,204,260	30,660,560
Engineering	12,271,040	12,271,040
Equipment + Tools During Construction	3,844,930	3,844,930
Contingency	13,312,720	12,858,350
Profit + Insurance	17,572,785	16,973,020
DIRECT CONSTRUCTION COST	$164,012,685	$158,414,850

APPENDIX D

DESCRIPTION OF GAS TURBINE COST MODEL

The gas turbine engines to be incorporated in advanced power stations similar to those identified in this study will be based on design philosophies different from those engines currently in production for present industrial applications. Thus, it was necessary to revise and enlarge a computer procedure previously developed(3) to estimate manufacturing costs of future gas turbine engines. Briefly, this manufacturing cost analysis program was based on information obtained directly from experts in the field of gas trubine design and from vendors who supply gas turbine manufacturers. A large number of individuals were consulted in an effort to obtain a broad data base of manufacturing cost information on a component-by-component basis. Correlations were then developed which related these costs to particular engine characteristics, e.g., physical dimensions, material selection, and production volume. After exhaustive examinations of less sophisticated approaches, it was estimated that the approach selected would provide the best correlations. In the resulting computer analysis the costs of all calculated components were summed, an allowance was made for the miscellaneous small engine parts, and assembly and test expenses were then added. Identifiable major component manufacturing costs were estimated to constitute over 80% of the total engine cost. The manufacturing cost computer program described was based on vendor contacts and cost correlations represented in 1970 dollar values.

A separate, Corporate-funded study was undertaken to revise the former cost correlation equations and to update the applicability of the calculation procedure to the 1974-1975 time period. This approach involved investigating the changes in costs for each of the major components during the intervening four year period. The data obtained were then separated into purchased part, raw material, and labor index values and were applied where appropriate to the cost correlations. This approach generally made it unnecessary to alter any of the original cost correlations, and program flexibility was accordingly preserved. By exercising the resulting proprietary computer program with input data representative of the engines identified during the study described in this report, it was then possible to estimate directly new engine costs which reflect the latest cost information available from a wide range of knowledgeable sources.

APPENDIX E

COGAS PERFORMANCE EVALUATION

The evaluation of COGAS performance estimates to identify inefficiencies associated with the various system components requires the development of an energy accounting system. In a combined gas and steam turbine cycle it is important that process heat requirements be satisfied by using gas turbine exhaust heat rather than by burning fuel or extracting heat from hot fuel gases prior to their use in the gas turbine. In this manner, the power generating capacity lost by the extraction of the process heat is only charged at steam cycle efficiency. The burning of fuel or extraction of heat from hot fuel gas results in a power loss that is charged at the full combined cycle efficiency. Therefore, a simple but effective approach to energy bookkeeping is to reflect fuel and heat consumption of individual components as an effective loss in fuel energy. Thus consumption of fuel gas or extraction of heat from the fuel gas prior to its use in the gas turbine is charged at the full energy value. However, the use of heat from the turbine exhaust is charged only at the ratio of steam cycle to combined cycle efficiency.

To determine the ratio of these efficiencies it is necessary to look at the system configured in Figure 61. In that figure, the gasifier air supply is separated from the main gas turbine flow stream in order to account for the heat of compression which is input to the gasifier. The equations that represent the system are:

POWER GENERATION

P_{GT} = Gas Turbine Output = $\eta_{GTLHV}\ HVR\ Q_{HHV}$ (49)

P_{STM} = Steam Cycle Output = $\eta_{STM}\ \eta_{BLR}\ Q_{EXH}$ (50)

$Q_{EXH} = HVR\ Q_{HHV} - \eta_{GTLHV}\ HVR\ Q_{HHV}$ (51)

Electrical Output $- (P_{GT} - P_{COMP} + P_{STM})\ \eta_{GEN}$ (52)

MECHANICAL EFFICIENCY

$$\eta_{cc(HHV)} = \eta_{GTHHV} + \eta_{STM}\,\eta_{BLR}\,HVR - \eta_{GTHHV} \qquad (53)$$

NOMENCLATURE

η_{GT} = Mechanical Efficiency of Gas Turbine (Output includes P_{COMP})

HVR = Ratio of Total Energy in Fuel LHV/HHV

Q_{HHV} = Total Chemical Plus Sensible Fuel Energy (HHV) (Above 59°F)

η_{STM} = Net Steam Cycle Mechanical Efficiency

η_{BLR} = Boiler Heat Removal Efficiency (Above 59°F)

Q_{EXH} = Heat in Gas Turbine Exhaust (Above 59°F)

P_{COMP} = Compressor Power for Gasifier Air Supply

η_{cc} = Combined Cycle Mechanical Efficiency

η_{GEN} = Generator Efficiency

These differ from the usual distillate representation in that they include the power required for gasifier air compression and are based on the higher heating value of the fuel. This latter fact requires the addition of a term that is simply the ratio of lower to higher heating value of the fuel at its supply temperature.